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16. Abstracts The economic and technical feasibility were studied of the adaptation of the CO_2 Acceptor Process to the problem of producing low-sulfur, low-Btu power plant fuel by the gasification of bituminous coals. Combustion of fuel gas makes possible a drastic reduction in NO_x emissions, compared with conventional PF stations equipped with stack gas cleanup processes. Atmospheric versus pressure operation of the gasifier were compared for the objective of preparing low-sulfur boiler fuel gas for an existing boiler. Capital and operating costs were estimated for two conceptual variations of the CO_2 Acceptor Process operated under 15 atm pressure, and for one lime bed gasification system operated at atmospheric pressure. The product gas is delivered to the boiler in both cases at the same pressure level of 25.7 psia. Pressure operation is preferred for this operation. Capital and operating costs were estimated for one pressurized variation of the CO_2 Acceptor Process integrated into two different conceptual power cycles: a super-charged boiler ahead of the gas turbine with an inlet temperature of 1800 F; and a high-temperature (2800 F) gas turbine with a heat recovery boiler on the turbine exhaust. A potential improvement in power cost is indicated with both power cycles, but the potential gain is much greater with the more advanced high temperature gas turbine cycle. Pretreatment studies were also conducted. The main reactor was used initially to study continuous preoxidation. Two coals were used which represent the extremes in caking properties among the high sulfur bituminous coals in the eastern half of the United States. Both coals were used in the 24 x 100 mesh size range. A simple laboratory test has been developed which provides a good screening procedure for testing the operability in the gasifier of preoxidized coals. Gasification studies were made in the continuous unit of operability at 1700 F not only with respect to caking but also of potential ash slagging due to insufficient rate of dissipation of the heat of combustion of the coal with incoming air. The experimental gasifier was operated at conditions simulating the full range of conditions in an envisioned full-scale unit. A limited amount of data were obtained from the gasification tests and in the demonstration run on the question of gasification kinetics. A single demonstration run was carried out using Disco char, a precarbonized coal char, as feedstock in which integrated operation of the gasification and regeneration steps were performed. The run encompassed circulation of acceptor through the two process steps. The purpose was to demonstrate the main features of the process and to delineate any special problem areas.

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DEVELOPMENT OF THE CO_2 ACCEPTOR PROCESS DIRECTED TOWARDS LOW-SULFUR BOILER FUEL

ANNUAL REPORT TO

Control Systems Division
Office of Air Programs
Environmental Protection Agency

under

OAP Contract No. EHSD-71-15

September 1, 1970 - November 1, 1971

by

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I. SUMMARY

A. Feasibility Study

The feasibility study called for in Phase I of this contract was directed at determining the economic and technical feasibility of the adaptation of the CO_2 Acceptor Process to the problem of producing low-sulfur, low-Btu power plant fuel by the gasification of bituminous coals. Combustion of fuel gas makes possible a drastic reduction in NO_x emissions, compared with conventional PF stations equipped with stack gas cleanup processes.

The first part of the feasibility study was directed at comparing atmospheric versus pressure operation of the gasifier for the objective of preparing low-sulfur boiler fuel gas for an existing boiler. Capital and operating costs were estimated for two conceptual variations of the CO_2 Acceptor Process operated under 15 atm pressure, and for one lime bed gasification system operated at atmospheric pressure.

All economic figures presented here are based on 1976 operation and include escalation at 7-1/2% per year on materials and labor and interest at 7-1/2% per year during construction.

The product gas is delivered to the boiler in both cases at the same pressure level of 25.7 psia. The results of this part of the feasibility study are summarized in Table I, comparing one of the pressurized cases with the atmospheric case. Figure 1 shows the impact of coal cost on the cost of low-sulfur boiler fuel for both cases. It is clear that pressure operation is preferred for this operation.

These calculations indicate that pressure operation not only produces a cheaper fuel gas but also effects a more efficient overall sulfur removal. The sulfur removal efficiency can be as high as 96% from a coal containing 4.3% sulfur. The sulfur in the gas is thus equivalent to that in a 0.3% sulfur fuel oil.

The above comparison was based on Case I the lower cost of the two pressure cases studied. The two pressure cases differ from each other in the method of regeneration of the elemental sulfur. Both involve CO_2 and H_2S acceptor reaction in the gasification section and deliver an equally low-sulfur pressurized producer gas. The second pressure case (Case II) involves a more expensive sulfur recovery section which adds about five percent to the capital and operating cost. New experimental data which have been obtained since this report was written emphasizes that Case II is technically more feasible than Case I. Although Case II is slightly more expensive, the difference in costs is well in precision of our estimate. The final conclusions remain the same. The second pressure case is also more attractive economically than the low pressure case. The two pressure cases are described in detail in Section III.

The second part of the feasibility study was concerned with utilization of the high pressure clean producer gas in an improved combined cycle power plant. Capital and operating costs were estimated for the first pressurized variation of the CO_2 Acceptor Process, described above, integrated

TABLE 1

Evaluation of Consol CO_2 Acceptor Process
Pressure vs. Atmospheric Operation

Generation of Fuel Gas for Conventional Boiler
70% Plant Operating Factor = 6132 hr/yr

	Case I Pressure Operation 15 atm.	Case III Atmospheric Operation
Coal Required		
Lb/hr (dry basis)	1,087,600	
MM Btu/hr (HHV)	13,812	
Product Gas		
Mols/hr	234,980	158,833
MM Btu/hr (HHV)	10,287	10,078
MM Btu/hr (HHV + sensible heat)	11,466	11,472
Temperature, °F	665	980
Pressure, psia	25.7	
% Sulfur Cleanup	96	86
Plant Investment(1)	$112,700,000	$141,700,000
Annual Operating Costs (ex. coal)	$10,090,000/yr	$11,410,000/yr
Coal at 40¢/MM Btu	$33,890,000/yr	$33,890,000/yr
Capital Charges at 15%	$16,900,000/yr	$21,260,000/yr
Sulfur Credit at $15/LT	($1,830,000/yr)	($1,650,000/yr)
Power Credit at 9 mills/KWH	($9,970,000/yr)	
Net Expenses	$49,080,000/yr	$64,910,000/yr
Fuel Cost Delivered to Power Station, ¢/MM Btu (HHV)	77.8	105.0
Fuel Cost Delivered to Power Station, ¢/MM Btu (HHV + sensible heat)	69.8	92.2

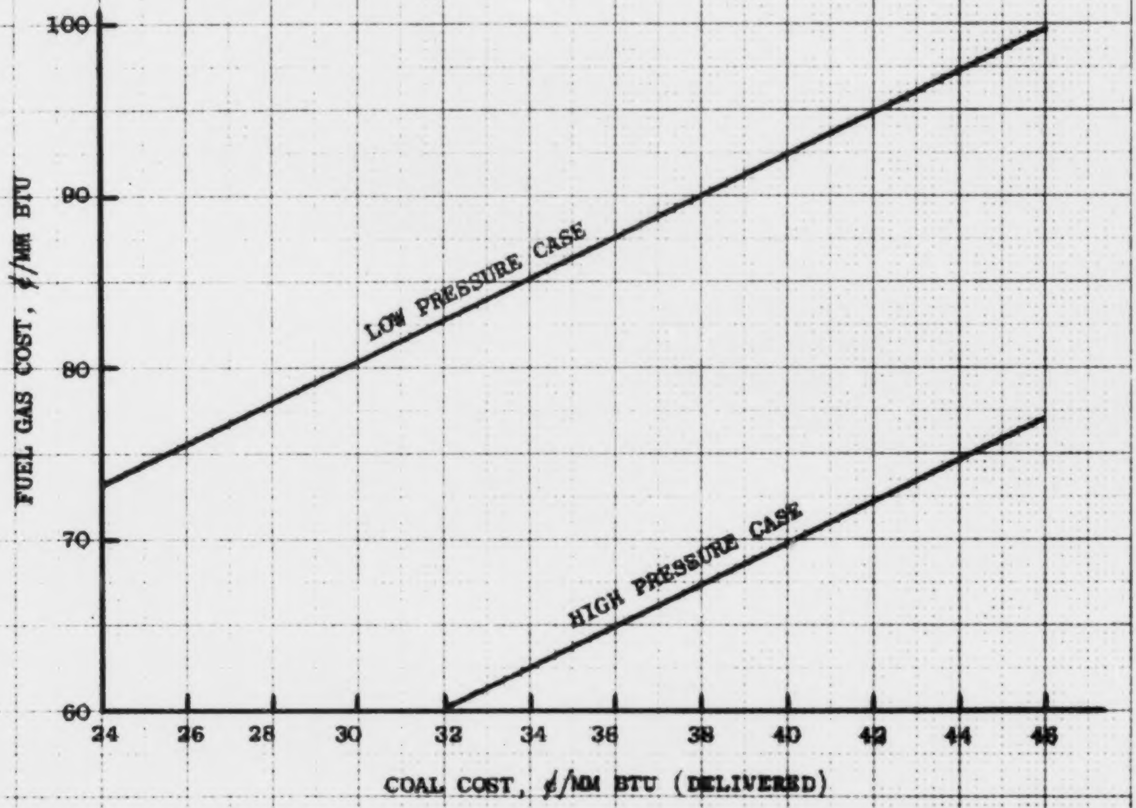
---

(1) 1976 operation, includes escalation and interest during construction.

FIGURE 1

FUEL GAS COST - EFFECT OF COAL COST

COST BASED ON HHV OF FUEL GAS
PLUS SENSIBLE HEAT CONTENT
1976 OPERATION



GRAPH PAPER

into two different conceptual power cycles. The first utilizes a supercharged boiler ahead of the gas turbine and is similar to the conceptual cycle now under study by Westinghouse Research Laboratories under Contract CPA-70-9 with the Office of Air Programs on fluidized combustion. This cycle utilizes currently available gas turbine technology with an inlet temperature of 1800°F.

The second power cycle uses the one proposed by the United Aircraft Research Laboratories (UARL) in their contract work with OAP on Advanced Power Cycles. It uses a high-temperature (2800°F) gas turbine with a heat recovery boiler on the turbine exhaust. Such a high-temperature turbine is well beyond the capabilities of current technology. UARL feels that the turbine technology required could be available for this cycle in the 1980's.

The investment, operating costs, and heat rates for the two combined cycles are compared with each other and with that of a conventional station equipped with SO_2 removal and sulfur recovery facilities in Table 2. A 15% capital charge was used for both conventional and future power stations.

Figure 2 shows the impact of coal cost on power costs for the different cases.

The values given for the two combined cycle cases in Table 2 and Figure 2, represent the base case for utilization of the Consol CO_2 Acceptor Process. The base case delivers clean hot producer gas at 1300°F to the power station. It is assumed that the gas can be cleaned sufficiently well by use of high pressure drop cyclones to permit its use in the combined cycle power plants. A more conservative case wherein the gas is cleaned by wet scrubbing is discussed in Section III.

In drawing conclusions from the figures presented in Table 2 and Figure 2, it should be recognized that the investment figures are only approximate. The bases upon which they are derived are likewise given in Section III. The figures given for overall station heat rates are, however, relatively precise. It should also be pointed out that economic optimization of the operation of the gasification system or its integration with the power cycle was not attempted. It was felt that optimization was not warranted until more experimental data were obtained at conditions required for the modified CO_2 Acceptor Process.

The numbers in Table 2 and Figure 2 illustrate the potential of the CO_2 Acceptor Gasification Process in producing a higher degree of sulfur removal, i.e., 96%, more efficient use of coal for power generation and power at a lower cost as compared with a conventional station equipped with stack gas scrubbing facilities.

The figures indicate a potential improvement in power cost with both power cycles considered, but the potential gain is much greater with the more advanced UARL gas turbine cycle.

In Table 2 the net power outputs of the combined cycle cases are considerably greater than that for the conventional PF station. As shown in Section III (Feasibility Study), multiple gasification process units are used. Therefore, the investment/KW would be little affected by reducing the output of the entire combined cycle power station.

TABLE 2

Summary of Power Costs and Investment

1976 Operation. See page 48 for plant cost basis.
70% Plant Operating Factor

System	Conventional Coal-Fired With Stack Gas Scrubbing	Case I - Gasification With Supercharged Boiler	Case I - Gasification With Exhaust Gas Cycle With 2800°F Turbine Inlet Temp.
Busbar Power, MW	1000(1)	1562	2053
Net Power, MW	1000	1544	1867
% Sulfur Removal (as elemental sulfur)	90	96	96
Overall Station Heat Rate, Btu/Net KWH	8900(2)	8930	7390
% Busbar Power From			
Steam Cycle	100	67.2	35.3
Gas Cycle	0	32.8	64.7
System Investment, $ KW Net Power			
Power Station	218.4	149.2	134.8
Sulfur Removal(3)	31.1	--	--
Fuel Gas Production Plant	--	61.1	60.2
Total Station	249.5	210.3	195.0
Power Cost	← Mills/KWH Busbar Power →		
Coal at 40¢/MM Btu	3.56	3.54	2.69
Gasification			
Direct Operating Cost	.77(3)	.94	.78
Capital Charges at 15% Inv.	.76(3)	1.48	1.34
Power Cost for Air Compression	--	--	.74
Power Plant			
Direct Operating Cost	.48	.48	.59
Capital Charges at 15% Inv.	5.34	3.61	3.00
Sulfur Credit at $15/LT	-.15	-.17	-.13
Total	10.76	9.88	9.01

(1) After subtraction of power for auxiliaries.

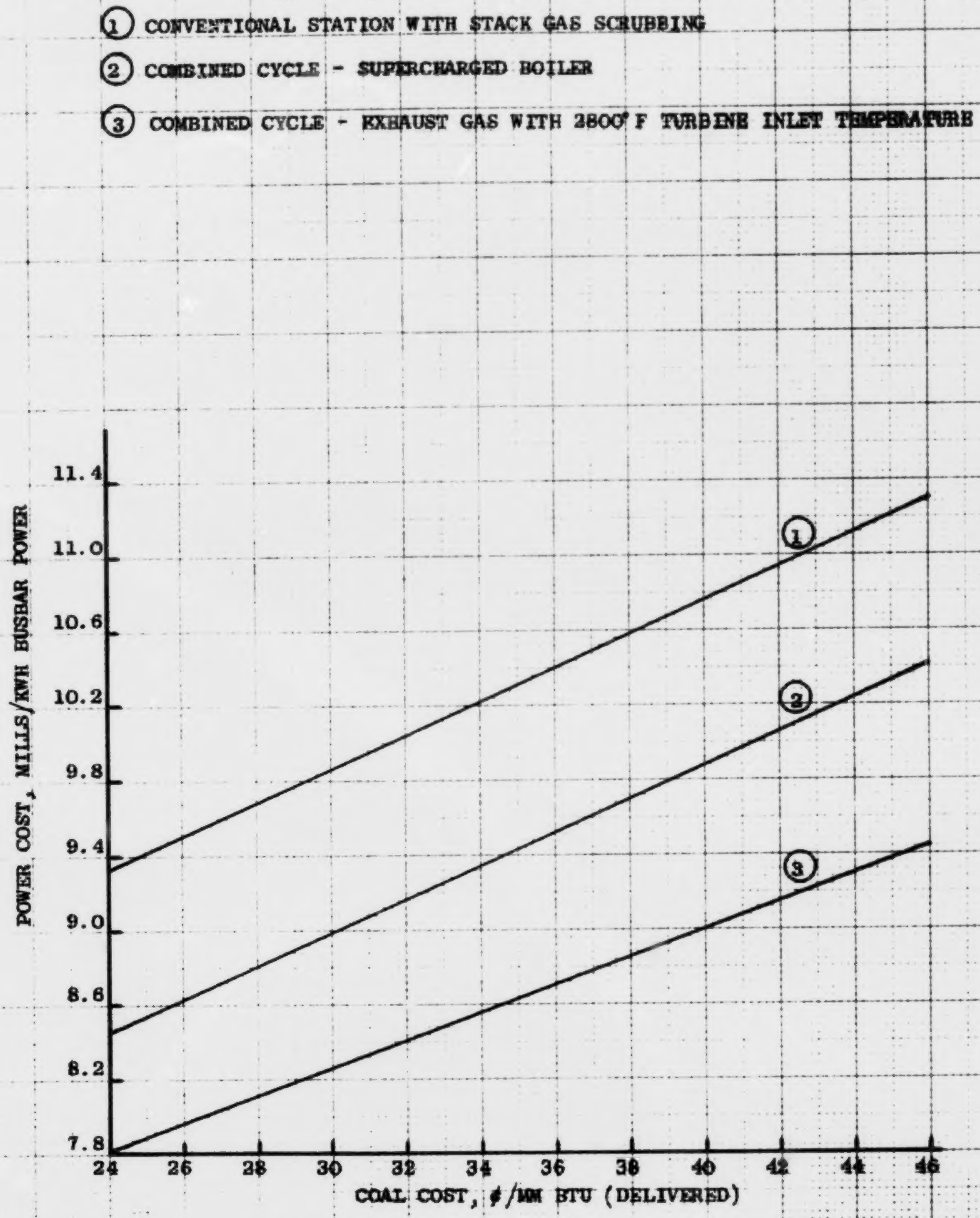
(2) Heat rate is 8900, based on coal burned in boiler. 640 Btu/KWH as fuel equivalent to generate reductant for sulfur recovery and power and steam requirements to drive process equipment.

(3) SO_2 removal and sulfur recovery plant via Consol's Formate Process. Fuel cost is included in direct operating cost at 80¢/MM Btu as purchased natural gas. Investment does not include cost of separate scrubber to remove residual particulate matter prior to SO_2 removal which may be required.

FIGURE 2

POWER COST

EFFECT OF COAL COST - 1976 OPERATION



B. Pretreatment Studies

The continuous unit which was used for the experimental study of the modified CO_2 Acceptor Process in Phase II-A of the contract was reconstructed and put into operation.

The main reactor was used initially to study continuous preoxidation. The preoxidation study on the continuous unit was preceded by a literature search which is included in Appendix E and by a brief exploratory laboratory study of the preoxidation kinetics. These two companion studies did not elucidate quantitatively all of the features of the preoxidation kinetics. They did show, however, that the reaction is very rapid and that the preoxidation kinetics are thus not an important process problem. These findings were confirmed by subsequent runs in the continuous unit.

The continuous unit was operated primarily to define the limits of operability in the preoxidation step and the minimum conditions of severity of pretreatment required to establish operability in the subsequent gasification step. Two coals were used in this study which represent the extremes in caking properties among the high-sulfur bituminous coals in the eastern half of the United States. Both coals were used in the 24 x 100 mesh size range.

The first coal, an Illinois No. 6 coal from the Hillsboro Mine, represented the more weakly caking coals. It was shown that this coal required pretreatment to establish freedom from caking in the subsequent gasification step. Operability was, however, readily established in both the pretreatment and gasification steps within the framework of preoxidation severity prescribed in the feasibility study, i.e., 8.5 wt % preoxidation.*

The second coal was a Pittsburgh seam coal from the Ireland Mine and represents the most highly caking coals. Operability limits were roughed out for both the pretreatment step and the severity of pretreatment required to establish operability in subsequent gasification. Preoxidation is operable at 750°F but inoperable at 800°F. Two-stage preoxidation is operable, wherein the first stage is conducted at 750°F and the second stage is conducted at 800°F but the product is of very low density and thus not readily suitable for subsequent gasification.

The minimum severity of preoxidation required in the case of Ireland Mine coal to establish operability in the subsequent gasification greatly exceeds the 8.5% preoxidation level called for in the feasibility study. The lowest level of preoxidation found to date which satisfies the above objective is 28 wt %. Further work should reduce the required amount of preoxidation.

A simple laboratory test has been developed which provides a good screening procedure for testing the operability in the gasifier of preoxidized coals.

* Severity of preoxidation is defined as:
Wt. % preoxidation = 100(lb O_2 reacted)/lb dry feed coal.

C. Gasification Studies

Studies were made in the continuous unit of operability in the gasification step at 1700°F not only with respect to caking but also of potential ash slagging due to insufficient rate of dissipation of the heat of combustion of the coal with incoming air. The experimental gasifier was operated at conditions simulating the full range of conditions in an envisioned full-scale unit. No ash slagging was detected in any of the runs even at oxygen partial pressures up to 2.4 atmospheres.

A limited amount of data were obtained as a byproduct of the gasification tests performed here and in the demonstration run discussed below on the question of gasification kinetics. The data indicate that the average gasification rate will be roughly comparable to the value assumed in the feasibility study.

D. Integrated Operation

A single demonstration run was carried out during this contract period using Disco char, a precarbonized coal char, as feedstock in which integrated operation of the gasification and regeneration steps were performed. The run encompassed circulation of acceptor through the two process steps. The purpose of the run was to demonstrate the main features of the process and to delineate any special problem areas.

The duration of the run was limited to 13 calcining-recarbonation cycles. However, the main features of the process were demonstrated. These include efficient removal of H_2S from the gas by the acceptor in the gasifier, lack of ash slagging in the regenerator at temperatures up to 1860°F, and rejection of sulfur from the acceptor in the regenerator. Acceptor life as determined indicates that the equilibrium activity will be higher than that estimated in the feasibility study. The Case I sulfur recovery system was not, however, fully demonstrated.

The run showed that a high degree of desulfurization can be achieved in the product char from the gasifier. The organic sulfur almost completely removed and the residual sulfur was largely tied up as CaS in the char ash.

II. CONCLUSIONS AND RECOMMENDATIONS

A. Feasibility Study

The CO_2 Acceptor Process as modified to produce low-sulfur boiler fuel has potential in two different types of applications as listed below:

1. Production of low-sulfur, low-Btu fuel gas for an existing boiler.
2. Production of low-sulfur, low-Btu fuel gas for a new combined cycle power plant.

An offshoot of the first application is also technically feasible, i.e., wherein both low-sulfur producer gas and char are produced as boiler fuel. This case is now being considered as an extension of this contract.

It has been shown in the feasibility study that operation of the gasification system under pressure is advantageous for both applications as compared with atmospheric pressure operation.

The major incentive for development of the modified CO_2 Acceptor Process resides in the second application. The integrated process has the potential of producing power at a lower cost, with lower emissions of SO_2 and NO_x, and with a higher efficiency of utilization of coal than is possible in a conventional station equipped with stack gas treating facilities.

The higher efficiency of such a power cycle is of great national importance in view of increasing fuel costs and the need to conserve our fossil fuel reserves.

The first application noted has less scope in the long run but can have shorter term use. Full utilization of the second application cannot be expected until the 1980's while the first can be put into use within the decade of the 70's. The first application is limited to situations where for various reasons it is not desirable or possible to add stack gas clean-up facilities to existing boilers. The alternative here is the use of low-sulfur fuel oil (ca. 0.3 wt %) which is not presently available in significant quantities. The CO_2 Acceptor Process can under certain circumstances, i.e., at inland locations close to coal supplies, potentially compete economically with imported low-sulfur fuel oil. It should also be noted that this application also has a better potential for NO_x control than a conventional station burning either coal or fuel oil.

B. Pretreatment Studies

The coal feed to the gasification step must be pretreated to establish operability in this step, i.e., to eliminate defluidization problems due to agglomeration and caking. The pretreatment step studied here was via preoxidation under full system pressure as called for in the feasibility study. This process was studied because it comprises the most convenient and economic way of carrying out the pretreatment step.

Operability limitations exist in the pretreatment step itself which are reflected in the maximum temperature that can be used. The maximum operable preoxidation temperature in the case of the most highly caking Pittsburgh Seam coals was found to be in the range of 750-800°F. For the more weakly caking Illinois No. 6 coals, operability is good at 800°F and the upper limit was not established.

Operability in both the pretreater and the gasifier is adversely affected by operation at elevated pressure as is desirable from the economic point of view. Also, a coal pretreated at atmospheric pressure is more severely decaked than one pretreated to the same degree of severity under pressure.

For the Illinois coal, the 8.5% preoxidation level required for adiabatic operation of the preoxidizer at 800°F is adequate to prevent caking in the gasifier. However, for the Pittsburgh Seam coal, a much more severe level of preoxidation (~ 28%) is required to prevent caking.

The effectiveness of the pretreatment process for a given level of preoxidation in decaking the coal increases with increasing preoxidation temperature and with decreasing particle size. An upper limit on the preoxidation temperature that may be used, however, is posed not only by operability in the preoxidizer but also by the properties of the preoxidized product. The particle density of the preoxidized coal, for example, decreases with increasing preoxidation temperature and thus reaches a point where it becomes unsuitable as a gasifier feedstock. The particle density under comparable conditions is higher the smaller the size of the pretreated coal. It also increases as the oxygen partial pressure in the preoxidation increases.

The preoxidation rates under process conditions are very rapid and thus offer no operating problem. Sufficient kinetic data were not obtained to derive a quantitative correlation of the effect of the variables on rate. The qualitative trends were in line with expectations from the literature on the subject. One interesting observation was made, however, in that the rate tended to become independent of particle size as the preoxidation temperature was increased.

Further experimental work is required to define conditions where the severity of pretreatment for the most highly caking coals may be substantially reduced over the levels so far demonstrated. Some particular subjects which should be studied are listed below:

1. Effect of more rigorous control of particle size with particular reference to elimination of coarse sizes.

2. Effect of oxygen partial pressure on degree of decaking achieved at a given preoxidation level.

3. Study multi-stage preoxidation (one or more stages) with increasing temperature between stages.

4. Same as above with the first stage carried out at atmospheric pressure.

Process studies should also be carried out in entrained phase preoxidation with a rising temperature regime as well as on other pretreatment methods which do not involve preoxidation.

C. Gasification Studies

The gasification step of the process must not only be operable from the point of view of caking but also it must be free from slagging due to ash fusion. It was concluded from the experimental work carried out under this contract that the gasification step will be free of operating difficulty from either cause if a properly pretreated feedstock is used.

The gasification kinetics as determined from the limited work carried out to date is of the same order as used in the feasibility study. More comprehensive experimental data are required to delineate the effects of temperature, carbon burnoff, coal type and gas composition on the gasification rate. It is recommended that such data be obtained in future work.

D. Integrated Operation

The main features of the process were demonstrated in a single demonstration run. The most important conclusions from this run were that acceptor activity is high enough to efficiently remove H_2S from the product gases in the gasifier, the acceptor activity for CO_2 absorption and its rate of decline is within or superior to the values estimated in the feasibility study, and that ash slagging in the regenerator is not a problem at least at operating temperatures of 1860°F or below.

Although rejection of sulfur from the acceptor was demonstrated, all ramifications of sulfur recovery via the Case I process were not proven. In particular, operation at the CO levels postulated in the feasibility study was not achieved nor was the recovery of elemental sulfur in the predicted quantity proven.

Further work is recommended to obtain more quantitative data on the following questions:

1. Operating limitations in the regenerator such as temperature, etc., as dictated by ash slagging.

2. Equilibrium acceptor activity as a function of process conditions in integrated operation.

3. Limitations imposed on recovery of sulfur as SO_2 and elemental sulfur via the Case I route.

III. FEASIBILITY STUDY

A. Introduction

The process development described herein is a logical extension of the development of the CO_2 Acceptor Process now being carried out under the auspices of the Office of Coal Research. The Office of Coal Research work is directed towards the production of high-Btu gas from low-rank Western coals. A pilot plant to demonstrate the process is now nearing completion at Rapid City, South Dakota. The basic technology, process economics and prior bench-scale research on the process as applied to low-rank coals are described in detail in reports to the Office of Coal Research.(1)

The modified process now under development under the auspices of the Office of Air Programs (OAP) is aimed at the production of low-sulfur boiler fuel rather than high-Btu gas. Another major distinction is that the process is directed towards the utilization of high-sulfur Eastern bituminous coals rather than low-sulfur Western coals.

The boiler fuel is either low-sulfur producer gas or a mixed product, i.e., producer gas and low-sulfur char. These products may be utilized either in a conventional boiler or in a new combined cycle plant. Generally speaking, it would be desirable to utilize the producer gas in a combined cycle power plant. The integration of coal gasification processes with combined power generation is, of course, an old concept. As a matter of fact, a 170 MW power plant utilizing this principle and the Lurgi pressurized gasification process is now under construction in Germany. The plant has been described by Rudolph in a recent paper.(2) Similarly, United Aircraft(3) in a study conducted for OAP have discussed the use of a slagging entrained phase pressurized gasification process to supply gas to a gas turbine based advanced power cycle.

Where low-sulfur char is produced as a co-product, it could be used as fuel for a fluidized-bed boiler or could be burned along with the gas in a conventional boiler.

The modified process as originally conceived was described in a paper(4) presented before the Second International Conference on Fluidized Combustion. Some changes, of course, have been made in the process concept since initiation of work under this contract, and they are discussed later in this report.

The process differs from other gasification processes previously proposed in that it removes sulfur within the process proper thus obviating the need for sulfur removal from the product gas. The Lurgi gasification process does have the advantage over this as well as over all other gasification processes proposed in that it is fully developed and is available for commercial exploitation. The Lurgi process is, however, limited to the use of +1/8-inch size coal and to coals that are not too strongly coking. The modified CO_2 Acceptor Process under development for OAP should be free of these limitations.

The initial phase of the program was a process design study to determine economic and technical feasibility. The study included a comparison of atmospheric and pressurized versions of the gasification process. Integration of the gasification process with combined cycle power generation was considered in order to arrive at an understanding of potential merits of the system.

Simultaneously with the process design study, construction of experimental equipment to study key features and potential problem areas in the laboratory has been completed. The major process features involved are listed below. The order corresponds more or less to the time sequence of the experimental program.

1. Coal pretreatment and elimination of coking problems in the gasifier.
2. Methods of introduction of air into the gasifier to avoid ash fusion problems.
3. Methods of rejecting and recovery of sulfur from the acceptor.
4. Acceptor life and activity in cycling through process conditions peculiar to the modified process. As a corollary, methods of reconstituting spent acceptor will be studied.
5. Operating limitations imposed by ash fusion in the regenerator.
6. Kinetics of gasification of Eastern coals under required process conditions.
7. Kinetics of acceptor reactions with CO_2 and H_2S in the gasifier.

If the above potential problem areas are successfully resolved in the laboratory, then it will be possible to consider the next stage in development of the process. This is demonstration of the process at the Rapid City pilot plant after required modifications to the plant are made. The plant as now designed is dedicated to the production of high Btu gas from low-rank Western coals.

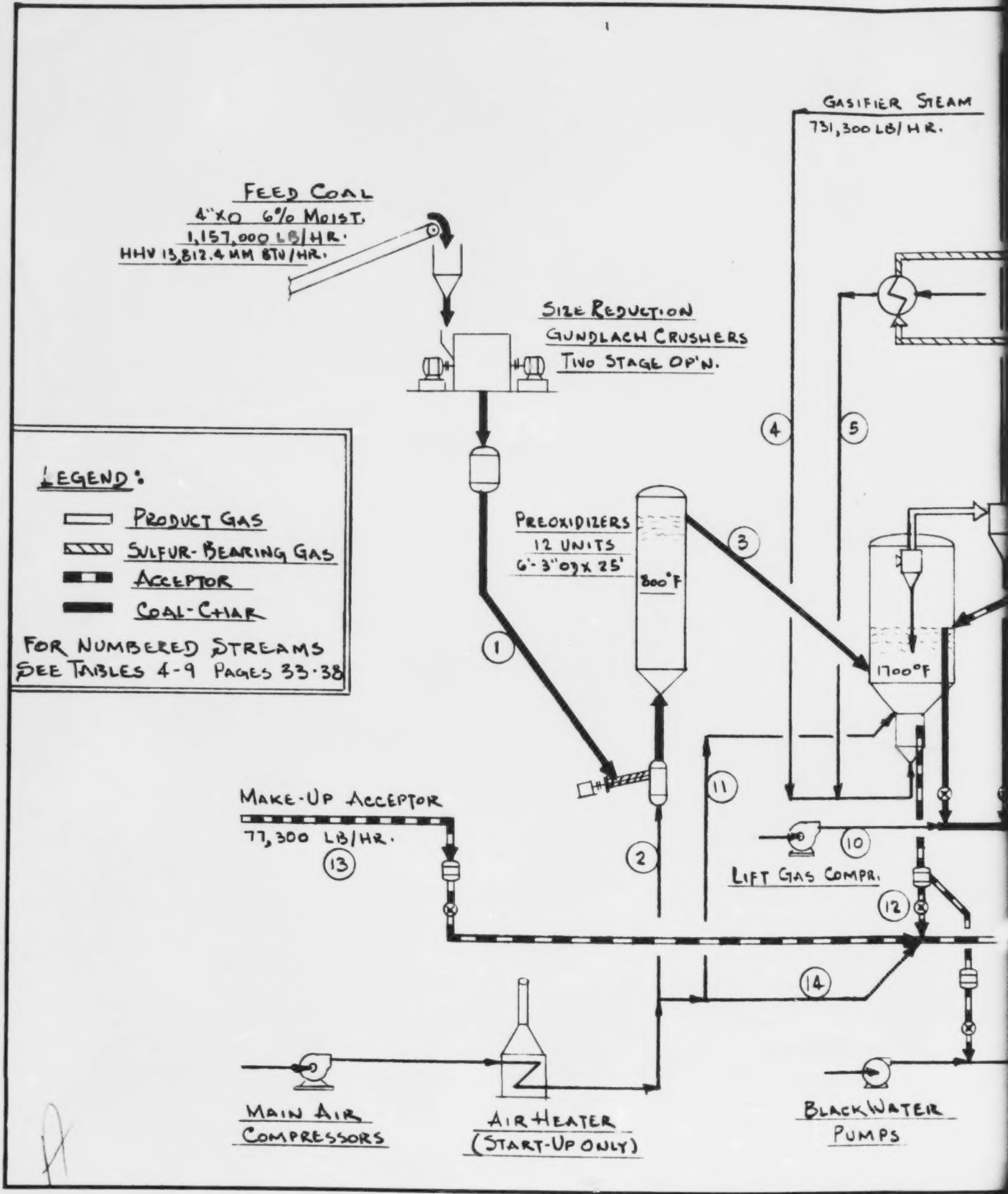
B. Feasibility Study - Phase I of Contract

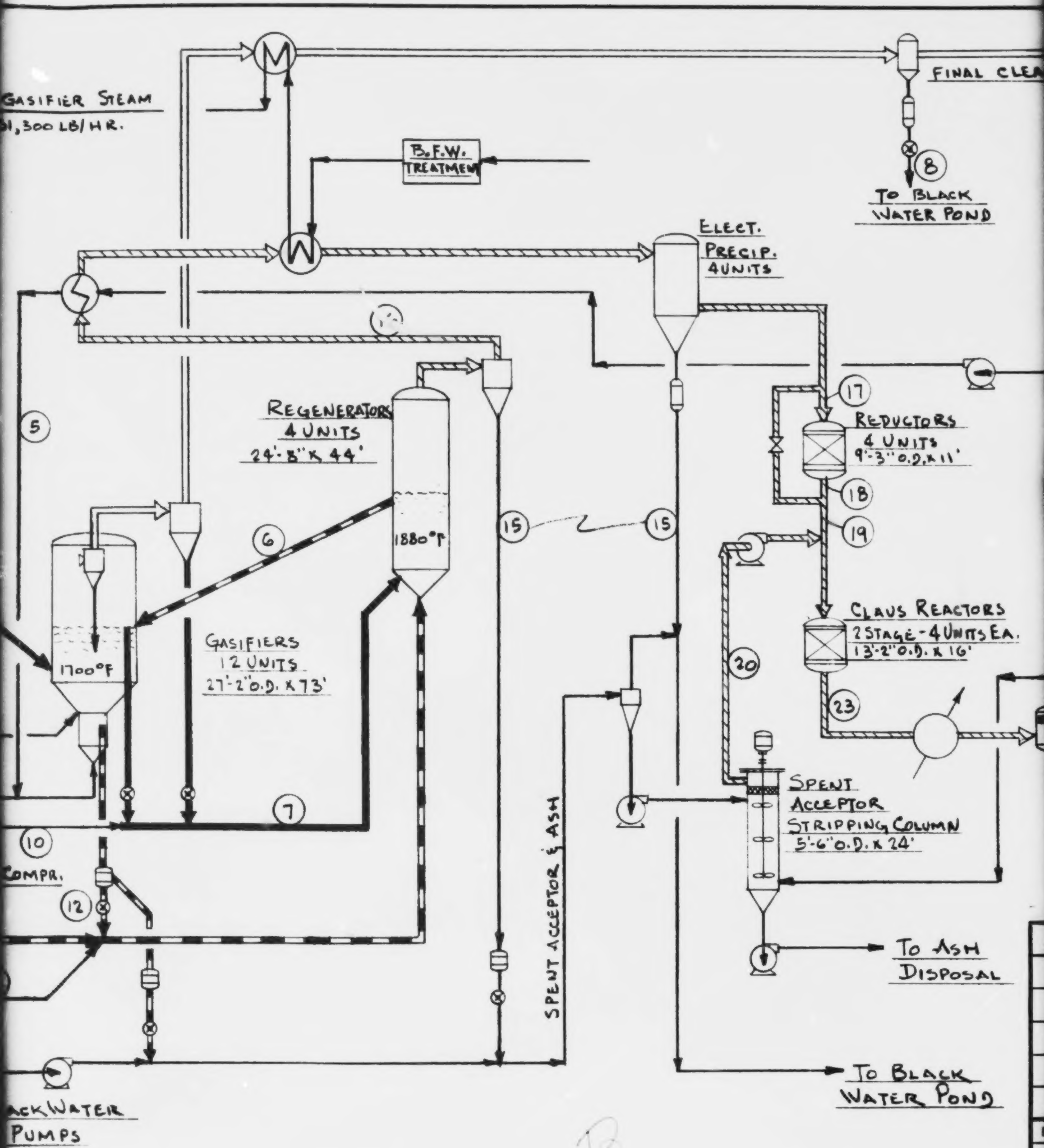
1. Process Description - Gasification Section

The size of the gasification plant was established to produce 1000 MW of electricity by conbustion of the cold "washed" product gas in a conventional boiler at a heat rate of 10,200 Btu/KWH. Since most of the cycles considered here generate electricity at a much higher efficiency, the net power output for the different cases varied and was generally much higher than 1000 MW, i.e., up to 1867 MW. Since multiple process units are used, the investment/KW would be little affected by reducing the size of the gasification plant.

Basic Pressure Case I (Dwg. XF-3143), Figure 3

The basic pressure case is shown schematically in Figure 3. Pittsburgh seam coal (12,884 tons/stream day) is charged to the gasification section of the plant by a system of belt and distributing conveyors, and is reduced in size from 4" x 0 to approximately an 8 Tyler mesh x 0 size consist. This size reduction is accomplished in two successive stages. Both stages employ multiple units of double-cage mill units to minimize excess fines production.


FEED COAL
4"x0 6% MOIST.
1,157,000 LB/HR.
HHV 13,812.4 MM BTU/HR.
SIZE REDUCTION
GUNDLACH CRUSHERS
TWO STAGE OP'N.
GASIFIER STEAM
731,300 LB/HR.
LEGEND:
▭ PRODUCT GAS
▨ SULFUR-BEARING GAS
ACCEPTOR
COAL-CHAR
FOR NUMBERED STREAMS
SEE TABLES 4-9 PAGES 33-38
PREOXIDIZERS
12 UNITS
6'-3" OD x 25'
800°F
1700°F
①
②
③
④
⑤
⑩
⑪
⑫
⑬
⑭
MAKE-UP ACCEPTOR
77,300 LB/HR.
LIFT GAS COMPR.
MAIN AIR
COMPRESSORS
AIR HEATER
(START-UP ONLY)
BLACKWATER
PUMPS



GASIFIER STEAM
1,300 LB/HR.
M
B.F.W. TREATMENT
W
FINAL CLEA
8
TO BLACK WATER POND
ELECT. PRECIP. 4 UNITS
5
REGENERATORS 4 UNITS 24'-8" x 44'
17
REDUCTORS 4 UNITS 9'-3" O.D. x 11'
18
19
6
1880°F
15
15
GASIFIERS 12 UNITS 27'-2" O.D. x 73'
1700°F
CLAUS REACTORS 2 STAGE - 4 UNITS EA. 13'-2" O.D. x 16'
20
23
SPENT ACCEPTOR STRIPPING COLUMN 5'-6" O.D. x 24'
7
10
COMPR.
12
SPENT ACCEPTOR & ASH
TO ASH DISPOSAL
TO BLACK WATER POND
ACKWATER PUMPS
B

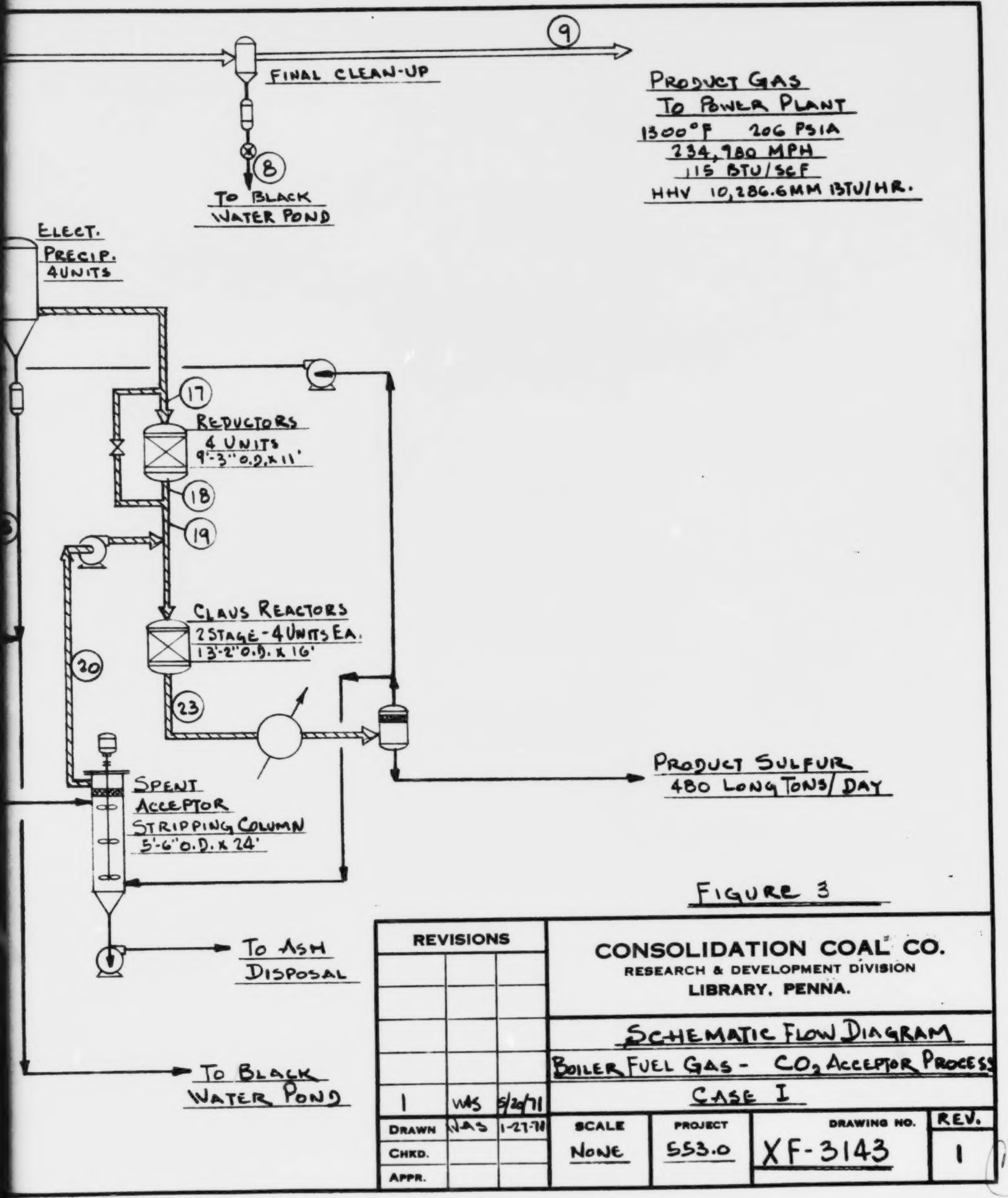

9
FINAL CLEAN-UP
8
TO BLACK WATER POND
PRODUCT GAS TO POWER PLANT
1300°F 206 PSIA
234,980 MPH
115 BTU/SCF
HHV 10,286.6MM BTU/HR.
ELECT. PRECIP. 4UNITS
17
REDUCTORS
4 UNITS
9'-3" O.D. x 11'
18
19
CLAUS REACTORS
2 STAGE - 4 UNITS EA.
13'-2" O.D. x 16'
20
23
PRODUCT SULFUR
480 LONG TONS/DAY
SPENT ACCEPTOR STRIPPING COLUMN
5'-6" O.D. x 24'
TO ASH DISPOSAL
TO BLACK WATER POND

FIGURE 3

CONSOLIDATION COAL CO.
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SCHEMATIC FLOW DIAGRAM
BOILER FUEL GAS - CO_2 ACCEPTOR PROCESS
CASE I

REVISIONS		
1	WAS	5/20/71
DRAWN	WAS	1-27-71
CHKD.		
APPR.		

SCALE	PROJECT	DRAWING NO.	REV.
NONE	553.0	XF-3143	1

The prepared coal is then lockhoppered into fluidized, air-blown preoxidizers which operate at approximately 800°F. Twelve parallel units are supplied. For this study, it is assumed that a preoxidation level of about 8.5 wt % oxidation is required with highly-caking Pittsburgh seam coals to maintain operability in the gasifiers. Experimental work under this contract will define the required level more precisely.

The total effluent from the preoxidizers is entrained overhead and passes to twelve parallel fluidized gasifiers together with additional air, steam, and tail gas from the Claus plant. Calcined acceptor from the regenerator flows in at the top of the gasifier bed, trickles down through the bed, and concentrates in a fluidized "boot" at the base. The main fluidized bed of the gasifier operates at 1700°F and 15 atm pressure. These operating conditions for the gasifier are dictated by the gasification rate of the fixed carbon in the coal, and by the kinetic and thermodynamic constraints of the CO_2 acceptor reaction. These constraints are discussed in greater detail in the section on Design Basis. Most of the sulfur in the coal reacts with the calcium oxide in the acceptor to form calcium sulfide. About 65% of the fixed carbon in the coal is gasified.

The remaining char from the gasifier bed is conveyed in a stream of inert gas to four parallel regenerators where it is burned with a deficiency of air at about 1880°F. Partially carbonated acceptor from the "boot" of the gasifier is also conveyed to the regenerator in an air stream. At the operating conditions of the regenerator, the calcium carbonate in the acceptor is calcined, and the calcium sulfide in the acceptor reacts to release sulfur dioxide and sulfur vapor. As noted, the calcined acceptor then flows back to the top of the gasifier bed completing the acceptor "loop". Of the heat required to maintain the gasifier at 1700°F, about 20% is supplied by the sensible heat in the calcined acceptor, about 30% by the chemical reaction of carbon dioxide with the calcium oxide in the acceptor, and the remaining 50% by combustion of a portion of the char with air.

Sulfur-bearing gas from the regenerator is cooled by preheating the Claus tail gas and by raising a portion of the required gasifier steam. After passing through electrostatic precipitators, approximately one-third of the gas flows to four parallel packed beds of alumina catalyst (called reductors) operated at a temperature of 800°F. In the reductors, practically all of the SO_2 in the gas is reduced to S_2, H_2S, and COS. The effluent gas from the reductor is then remixed with the by-passed regenerator offgas. The molal ratio of $(H_2S + COS)/SO_2$ is then exactly 2.0.

This stream is then fed to a modified two-stage Claus system where about 96% of the sulfur in the regenerator offgas is recovered as liquid sulfur. The tail gas from the Claus system is then preheated to 1200°F (by indirect heat exchange with the regenerator offgas) and charged to the gasifier. This CO_2-rich gas provides the partial pressure driving force for recarbonation of the acceptor and thereby supplies about 30% of the gasifier heat requirement, as noted above.

The product gas from the gasifier is cooled to about 1300°F by raising most of the required gasifier steam. In the simplest conception of this process, the gas (234,980 mols/hr of 115.5 Btu/SCF HHV) would then be delivered directly to the power station at a temperature of about 1300°F and at a pressure of about 206 psia. If later experience shows that the gas-fired turbine in the power station is too sensitive to any alkali content in the product gas, then the gas would readily be water-scrubbed in a venturi-type scrubber to the gas dew point of about 240°F and then reheated by indirect heat exchange with itself to a delivery temperature of about 1100°F and a pressure of about 200 psia.

Unburned char and char ash from the regenerator, and the rejected fraction of the acceptor are lockhoppered out of the system and water quenched in a circulating black water system. The solids in the black water circuit are concentrated by passage through conventional hydroclones, and stripped with a gas stream rich in carbon dioxide to convert the calcium sulfide in the acceptor to calcium carbonate. All of the rejected solids are then disposed of to an ash pond as in conventional wet-bottom, coal-fired boiler systems.

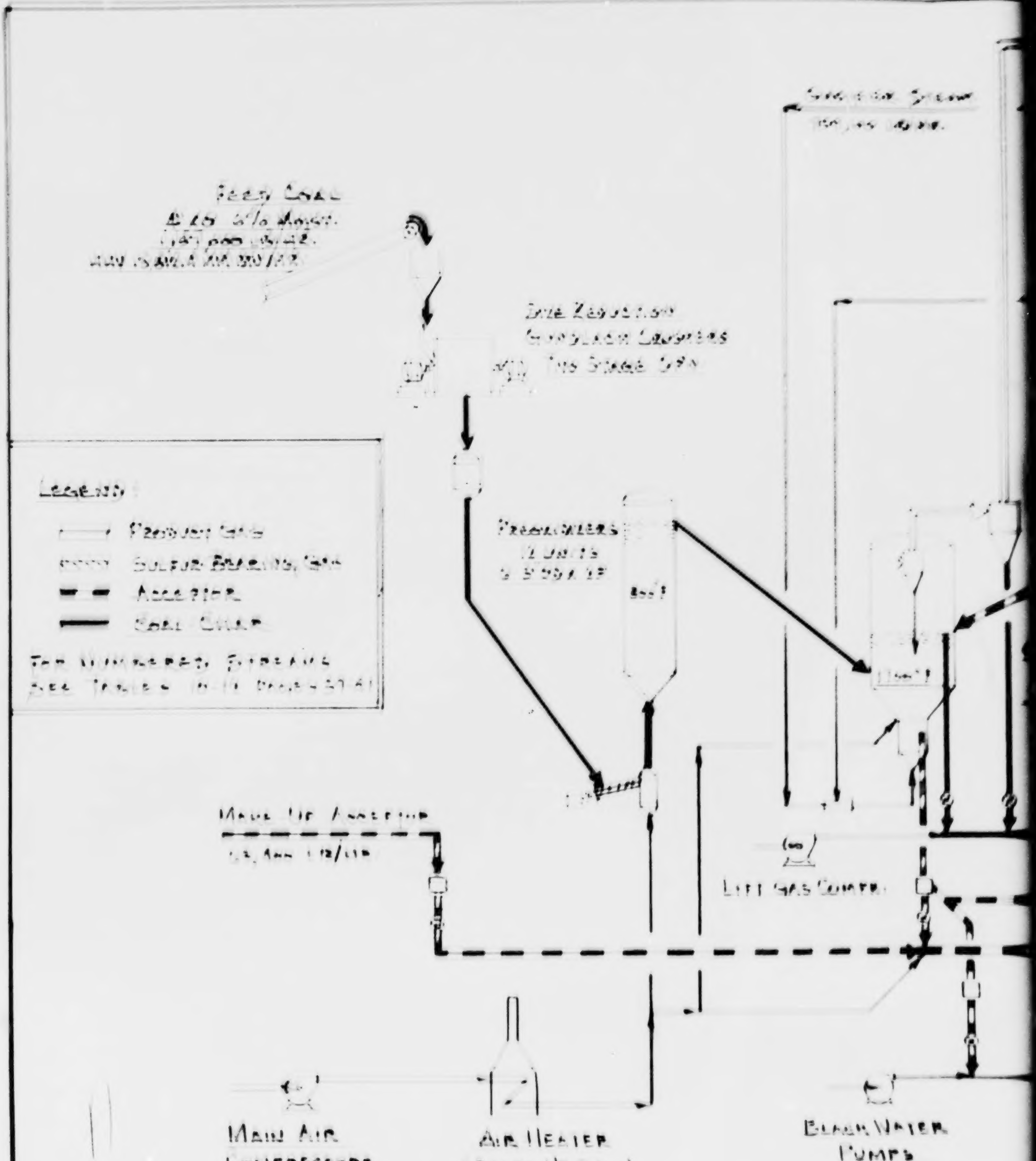
Alternate Pressure Case II (Dwg. XF-3144), Figure 4

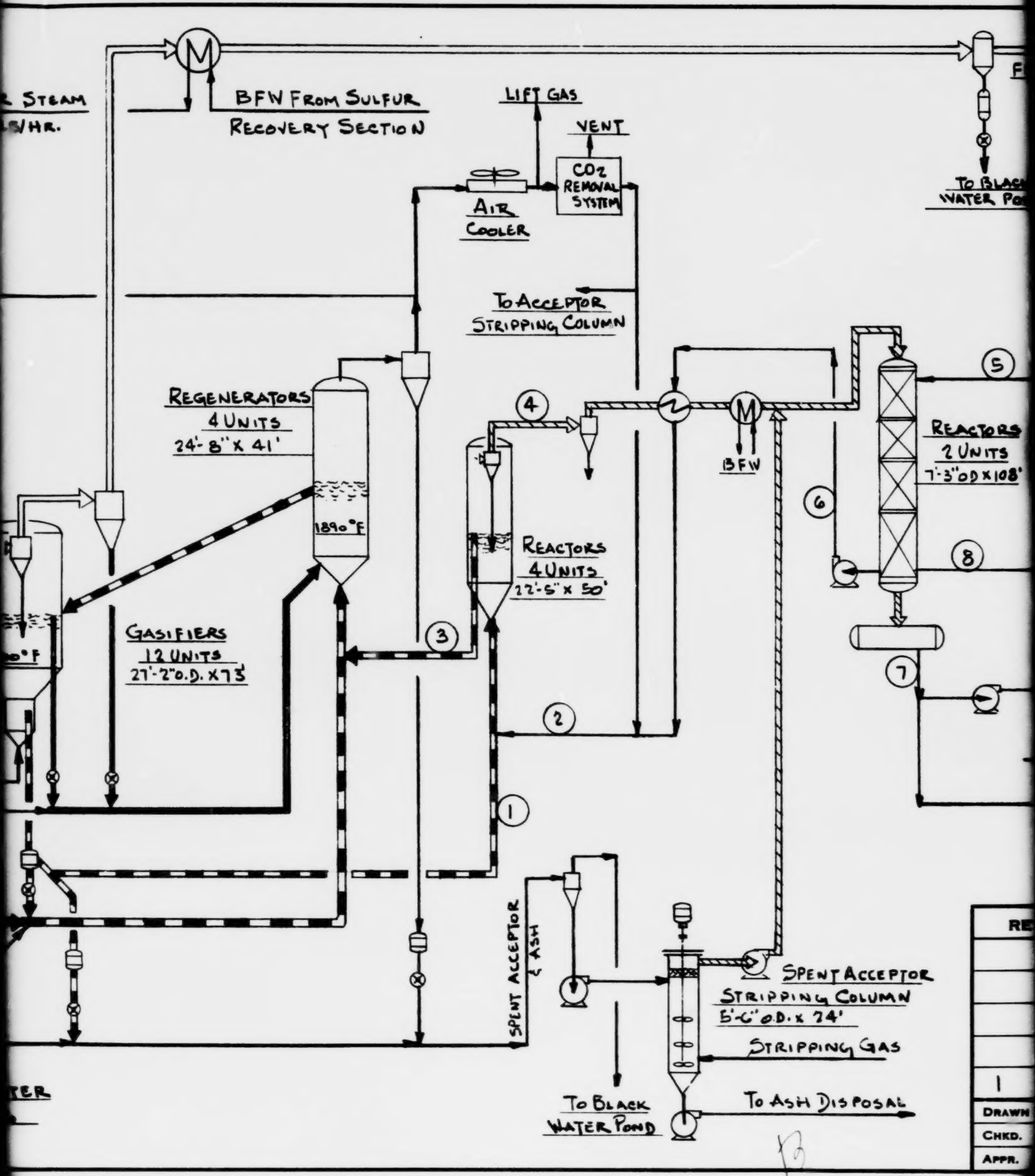
The alternate pressure case differs primarily in the method by which the sulfur is removed from the acceptor. In this case, the regenerator is operated with very slight excess air and the sulfur is not released from the acceptor. One-third of the circulating acceptor leaving the gasifiers is charged to a fluid bed of acceptor where it reacts at 1300°F with an equimolar mixture of CO_2 and water vapor to convert the calcium sulfide to calcium carbonate. The effluent gas from this reactor is then passed to a packed column where it mixes in a concurrent flow with a liquid stream of sulfurous acid. This reaction produces liquid sulfur at a temperature of 328°F and is a variation of the Wackenroder (5) reaction. Finally, a portion of the liquid sulfur (about one-third) is burned to produce the required sulfurous acid via an SO_2 absorption column.

In all other respects Cases I and II are quite similar.

Atmospheric Pressure Case III (Dwg. XF-3145), Figure 5

Under the terms of this OAP contract, the CO_2 Acceptor Process at pressure is to be compared with a similar case at atmospheric pressure. Under these conditions, there is no reaction of CO_2 with calcium oxide and the acceptor serves only as a sulfur acceptor and as a sensible heat carrier. Again, coal is prepared to 8 Tyler mesh by zero, preoxidized at 800°F and gasified at 1700°F with steam and air. A small amount of gasifier product gas is recycled to the bottom of the "boot" to prevent oxidation of the acceptor CaS by the incoming steam. The sensible heat in the gasifier gas is used to raise the required gasifier steam, and to raise the plant steam required to run the main air compressors. The final product gas (158,833 mols/hr of 167 Btu/SCF HHV) is then delivered to the power station at about 980°F and 25.7 psia. Again, if water scrubbing of the gas is required, it could be reheated to about 750°F.


Feed Coal
Legend
Product Gas
Sulfur Bearing Gas
Acceptor
Coal Char
Make Up Acceptor
Lift Gas Compr.
Main Air Compressors
Air Heater (Start Up Only)
Black Water Pumps



R STEAM
LB/HR.
BFW FROM SULFUR
RECOVERY SECTION
LIFT GAS
VENT
CO2 REMOVAL SYSTEM
AIR COOLER
F
TO BLAC
WATER PO
TO ACCEPTOR
STRIPPING COLUMN
REGENERATORS
4 UNITS
24'-8" X 41'
1890°F
4
BFW
5
REACTORS
2 UNITS
7'-3" OD X 108'
6
8
REACTORS
4 UNITS
22'-5" X 50'
GASIFIERS
12 UNITS
27'-2" O.D. X 75'
0°F
3
7
2
1
SPENT ACCEPTOR & ASH
SPENT ACCEPTOR
STRIPPING COLUMN
5'-6" O.D. X 24'
STRIPPING GAS
TER
TO BLACK
WATER POND
TO ASH DISPOSAL
RE
1
DRAWN
CHKD.
APPR.
B

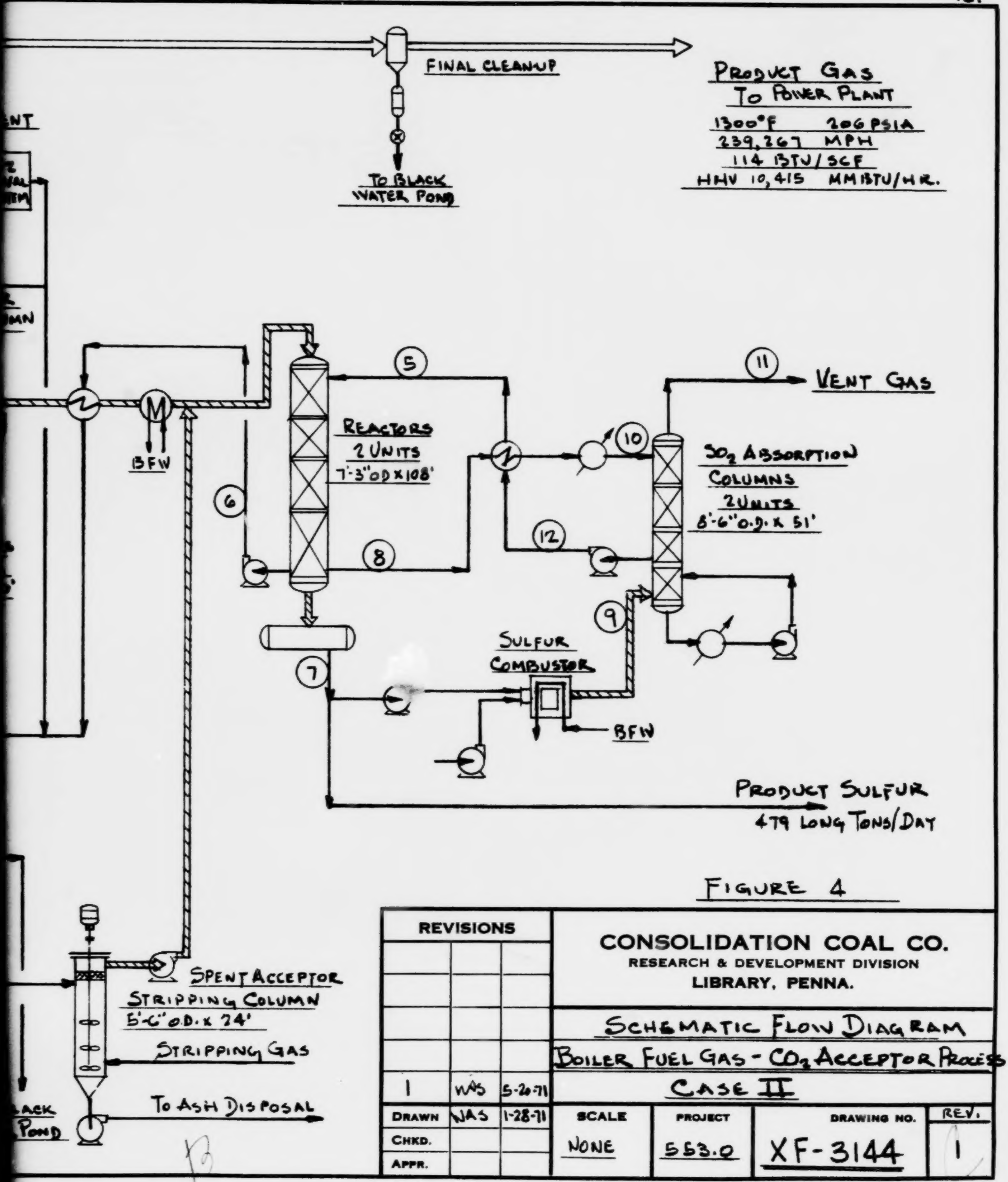

FINAL CLEANUP
TO BLACK WATER POND
PRODUCT GAS TO POWER PLANT
1300°F 206 PSIA
239,267 MPH
114 BTU/SCF
HHV 10,415 MMBTU/HR.
ENT
MN
BFW
REACTORS 2 UNITS 7'-3" OD x 108'
M
5
6
7
8
9
10
11
12
VENT GAS
SO$_2$ ABSORPTION COLUMNS 2 UNITS 8'-6" O.D. x 51'
SULFUR COMBUSTOR
BFW
PRODUCT SULFUR 479 LONG TONS/DAY
SPENT ACCEPTOR STRIPPING COLUMN 5'-6" O.D. x 74'
STRIPPING GAS
TO ASH DISPOSAL
LACK POND

FIGURE 4

REVISIONS		
1	WAS	5-20-71
DRAWN	WAS	1-28-71
CHKD.		
APPR.		

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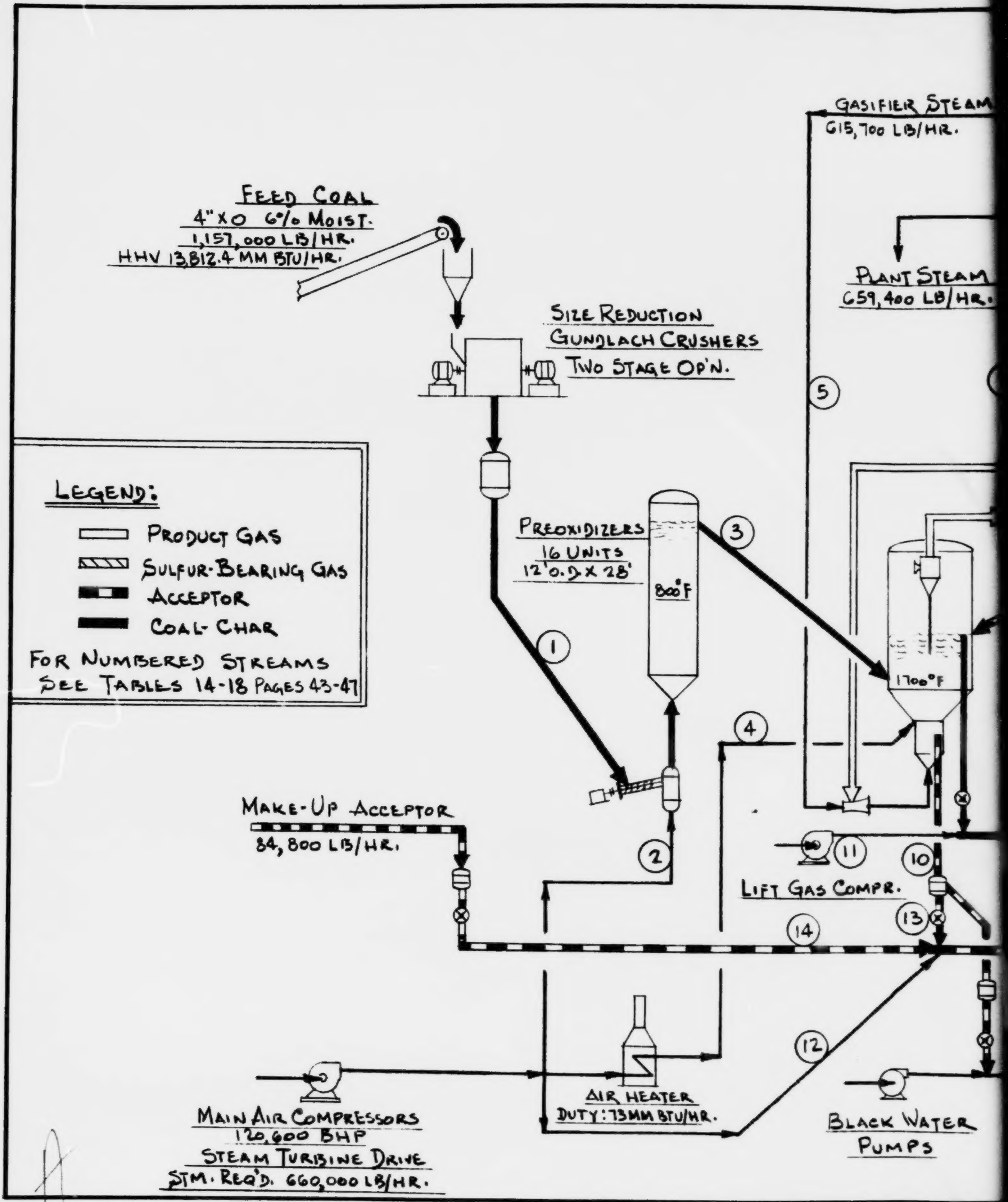
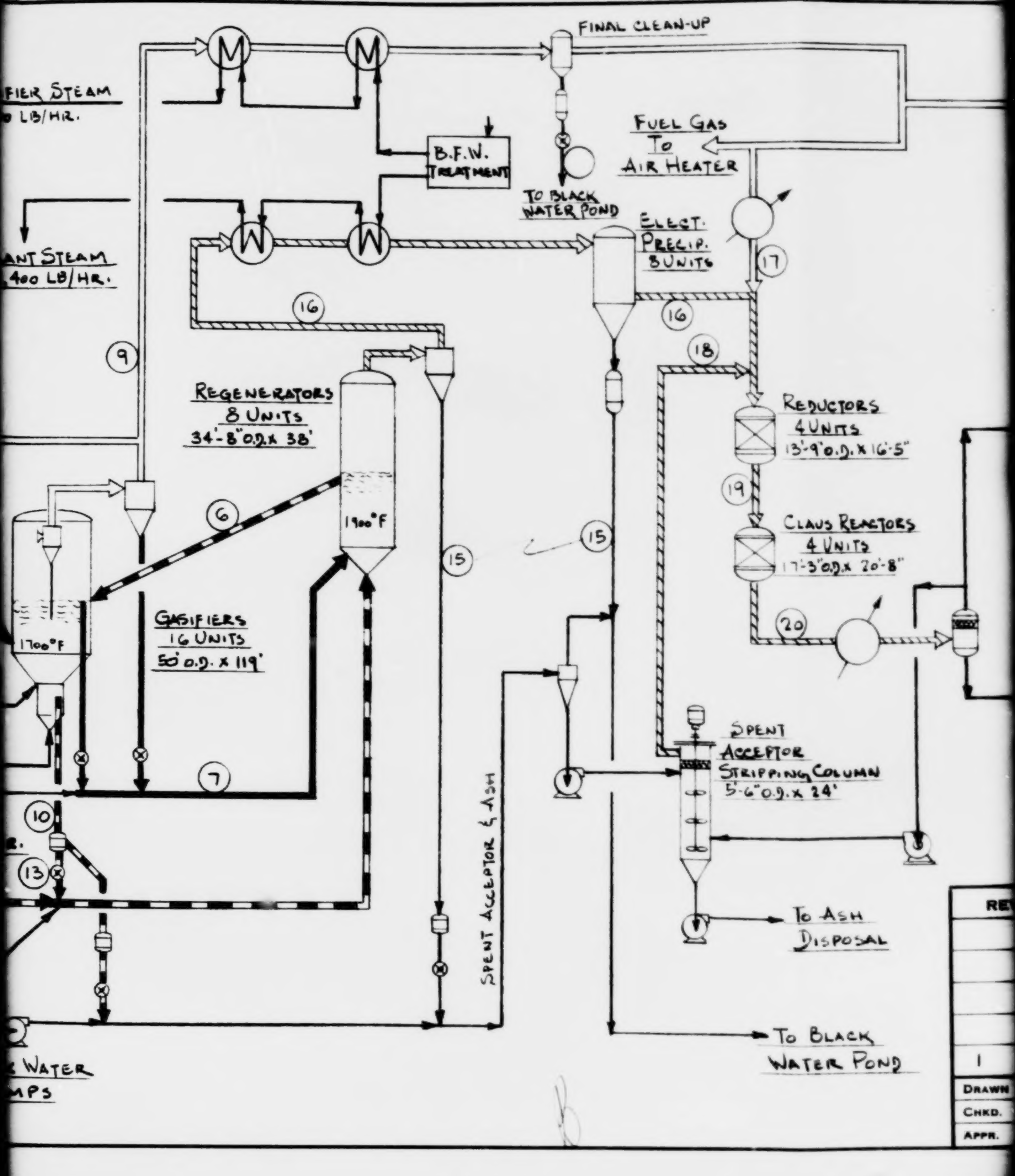
SCHEMATIC FLOW DIAGRAM
BOILER FUEL GAS - CO$_2$ ACCEPTOR PROCESS
CASE II

SCALE	PROJECT	DRAWING NO.	REV.
NONE	553.0	XF-3144	1


GASIFIER STEAM
615,700 LB/HR.
FEED COAL
4"×0 6% MOIST.
1,157,000 LB/HR.
HHV 13,812.4 MM BTU/HR.
PLANT STEAM
659,400 LB/HR.
SIZE REDUCTION
GUNDLACH CRUSHERS
TWO STAGE OP'N.
5
LEGEND:
PRODUCT GAS
SULFUR-BEARING GAS
ACCEPTOR
COAL-CHAR
FOR NUMBERED STREAMS
SEE TABLES 14-18 PAGES 43-47
PREOXIDIZERS
16 UNITS
12' O.D. × 28'
800°F
3
1700°F
1
4
MAKE-UP ACCEPTOR
84,800 LB/HR.
2
11
LIFT GAS COMPR.
10
13
14
12
AIR HEATER
DUTY: 73MM BTU/HR.
BLACK WATER
PUMPS
MAIN AIR COMPRESSORS
120,600 BHP
STEAM TURBINE DRIVE
STM. REQ'D. 660,000 LB/HR.


FINAL CLEAN-UP
FIER STEAM
0 LB/HR.
M
M
FUEL GAS
TO
AIR HEATER
B.F.W.
TREATMENT
TO BLACK
WATER POND
ANT STEAM
,400 LB/HR.
W
W
ELECT.
PRECIP.
8 UNITS
17
16
16
18
9
REGENERATORS
8 UNITS
34'-8" O.D. x 38'
REDUCTORS
4 UNITS
13'-9" O.D. x 16'-5"
19
6
1900°F
CLAUS REACTORS
4 UNITS
17'-3" O.D. x 20'-8"
15
15
1700°F
GASIFIERS
16 UNITS
50' O.D. x 119'
20
SPENT
ACCEPTOR
STRIPPING COLUMN
5'-6" O.D. x 24'
7
10
13
SPENT ACCEPTOR & ASH
TO ASH
DISPOSAL
TO BLACK
WATER POND
WATER
MPS
REV
1
DRAWN
CHKD.
APPR.

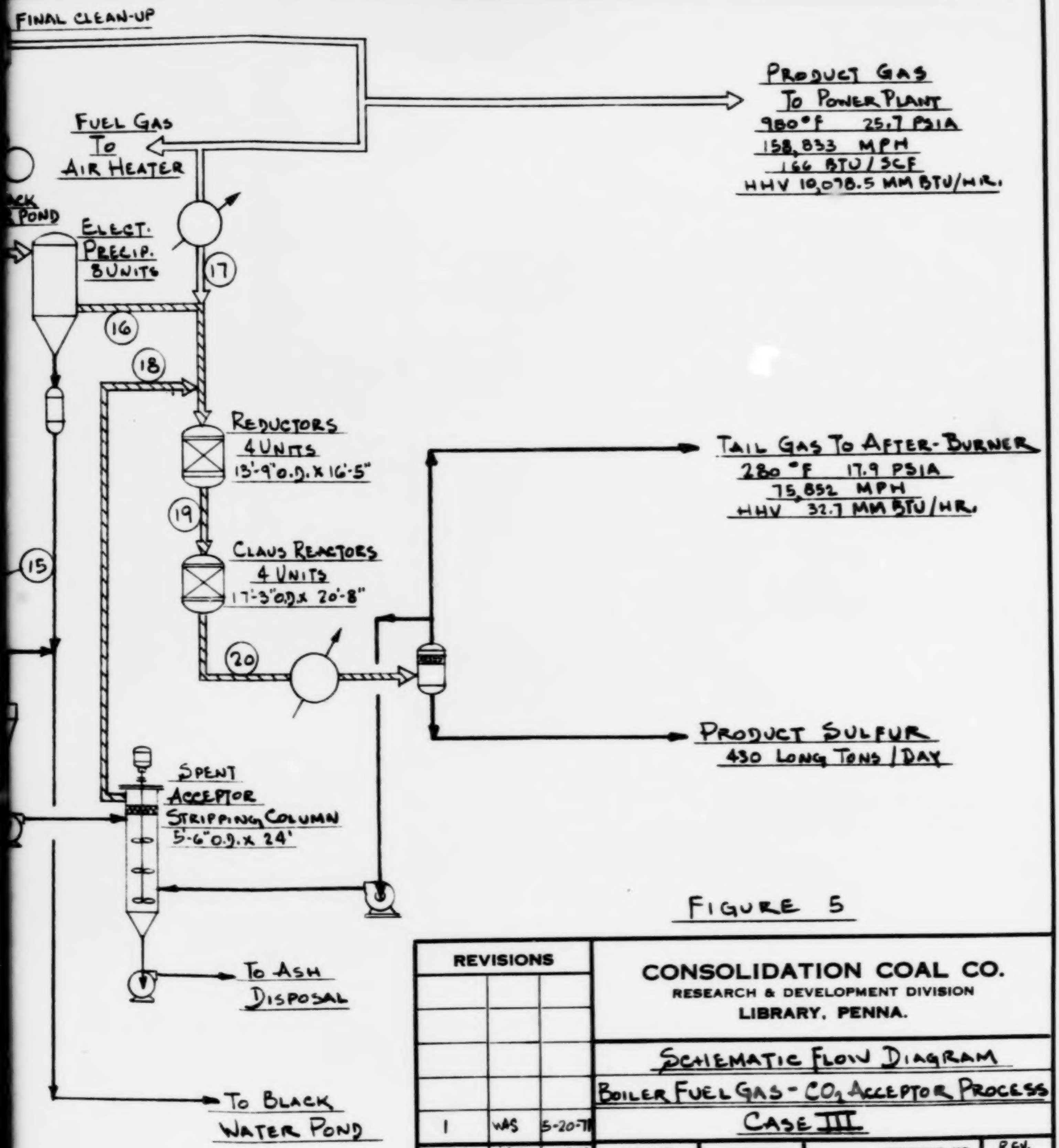

FINAL CLEAN-UP
PRODUCT GAS
TO POWER PLANT
980°F 25.7 PSIA
158,833 MPH
166 BTU/SCF
HHV 10,078.5 MM BTU/HR.
FUEL GAS
TO
AIR HEATER
ACK
POND
ELECT.
PRECIP.
8 UNITS
17
16
18
REDUCTORS
4 UNITS
13'-9" O.D. x 16'-5"
19
CLAUS REACTORS
4 UNITS
17'-3" O.D. x 20'-8"
20
15
TAIL GAS TO AFTER-BURNER
280°F 17.9 PSIA
75,852 MPH
HHV 32.7 MM BTU/HR.
PRODUCT SULFUR
430 LONG TONS/DAY
SPENT
ACCEPTOR
STRIPPING COLUMN
5'-6" O.D. x 24'
TO ASH
DISPOSAL
TO BLACK
WATER POND

FIGURE 5

REVISIONS		
1	WAS	5-20-71
DRAWN	WAS	1-29-71
CHKD.		
APPR.		

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SCHEMATIC FLOW DIAGRAM
BOILER FUEL GAS - CO_2 ACCEPTOR PROCESS
CASE III

SCALE	PROJECT	DRAWING NO.	REV.
NONE	553.0	XF-3145	1

The regenerator is operated at 1900°F with stoichiometric air and the calcium sulfide is converted to calcium oxide with the release of SO_2. The regenerator offgas is cooled by means of steam generation and the gas is passed to a catalyst-packed reductor and a one-stage Claus plant to convert and recover the sulfur as liquid sulfur. The Claus tail gas, still containing some H_2S, cannot be vented directly but has to be passed through the power station boiler fire box to convert the H_2S to SC_2.

As this is a feasibility study, optimization of each processing step by an extended study of alternates was not attempted. The following sections show the basis for design and the assumptions used for each major processing step. Some further details are given in Appendix B.

Feed Coal

This study is based on the use of a high-sulfur content, highly caking Pittsburgh seam coal. The analysis of a typical coal of this seam is shown below:

Feed Coal Analysis

Moisture, as received	6.0 Wt %
Ultimate Analysis, MF Basis	
Hydrogen	4.8 Wt %
Carbon	69.8 Wt %
Nitrogen	1.2 Wt %
Oxygen	7.6 Wt %
Sulfur	4.3 Wt %
Ash	12.3 Wt %
Higher Heating Value, MF Basis	12,700 Btu/lb

The coal is assumed to be delivered to the plant at run-of-mine size of 4" x 0.

Preoxidation

Highly caking Pittsburgh seam coal requires a pretreatment step (in this case, preoxidation) to maintain operability in an air-blown gasifier. The degree of preoxidation, and the temperature required, will be determined experimentally under Phase II-A of this contract. For this feasibility study, it was assumed that a preoxidation level of about 8.5 wt % at a temperature of 800°F would be adequate pretreatment. Operating conditions and yield structure are directly based on prior Consolidation Coal Company pilot plant data for atmospheric pressure operation.

The pertinent design data incorporated into this study are summarized below:

Preoxidizer Design Constraints

	Pressure Operation Cases I and II	Atmospheric Pressure Case III
Temperature, °F	← 800 →	
Outlet Pressure, psia	224	41
Percent Preoxidation(1)	8.6	8.8
Oxygen Utilization, %	← 78 →	
Partial Pressure Oxygen, psia		
Inlet	47.7	10.1
Outlet	7.6	1.4
Ln Mean	21.8	4.4
Outlet Fluidizing Velocity, ft/sec	← 1.25 →	
Avg. Solids Residence Time, min.	10	50

(1) 100 (lb O_2 consumed)/(lb MF coal).

As is obvious from the above, it was assumed that the rate of oxidation is directly related to the oxygen partial pressure.

Gasification

There are numerous design conditions and constraints that are of importance in fluidized, air-blown gasification of a caking coal. Consol has been actively engaged in research in this field, and in the use of an active acceptor to supply heat to the gasifier for over twenty years. The following compilation of bases and assumptions was drawn from this background information.

Gasifier Design Basis

	Pressure Operation Cases I and II	Atmospheric Pressure Case III
Temperature, °F	← 1700 →	
Outlet Pressure, psia	220	35
Outlet Fluidizing Velocity, ft/sec	← 1.09 →	
Acceptor		
Type	← Dolomite →	
Wt % Impurities	← 15 →	
Activity(1)	.13	--
Make-up Rate(2), %	0.5	0.2
Avg. Residence Time in Bed, min.	← 7 →	
% Fixed Carbon Gasified	← 65(3) →	
% Hydrogen in Residual Char	← 0.4 →	
Outlet Gas Composition		
Partial Pressure CO_2, atm	1.35	0.14
Partial Pressure H_2O (v), atm	1.81	0.23
Shift Equilibrium	← Yes →	
Equilibrium in $CaO + H_2S = CaS + H_2O$	← Yes →	
Steam Conversion, %	37	58
Mean Gasification Rate, Atoms fixed C Gasified/atom C in bed/min.	75×10^{-4}	15×10^{-4}

(1) (mols $MgO \cdot CaCO_3$ formed in gasifier)/(mols $MgO \cdot CaO$ fed to gasifier).

(2) 100 (mols/hr $MgCO_3 \cdot CaCO_3$ added)/(mols/hr $MgO \cdot CaO$ circulated).

(3) Approximately 80% of total carbon fed.

Regeneration

Regenerator Design Conditions

	Pressure Operation Cases I and II	Atmospheric Pressure Case III
Temperature, °F	1883	1900
Outlet Pressure, psia	220	35
Outlet Fluidizing Velocity, ft/sec	1.6	2.5
% Carbon Burnout	← 90 →	
Avg. Acceptor Residence Time, min.	10	20
Outlet Gas Composition		
Partial Pressure CO, atm	0.38	0
Partial Pressure CO_2, atm	4.4	0.44
Shift Equilibrium	← Yes →	

2. Process Description and Design Basis - Sulfur Removal and Recovery

This section gives a more complete description of the sulfur removal and recovery features than has been discussed above. The design basis is also given in more detail.

Sulfur Removal

In Eastern steam coals, sulfur occurs mostly in the forms of iron pyrites and organic sulfur. In all three Cases included in this study the coal sulfur undergoes the following reactions in the gasifier, operated at partial-combustion conditions:

$$FeS_2 = FeS + 1/2\ S_2$$

$$[S]_{org} + O_2 = SO_2$$

$$SO_2 + 2\ H_2 = 2\ H_2O + 1/2\ S_2$$

$$FeS + H_2 = Fe + H_2S$$

$$[S]_{org} + H_2 = H_2S$$

$$1/2\ S_2 + H_2 = H_2S$$

All the above reactions are favorable, at the conditions used, with respect to kinetics and equilibrium, with the result that substantially all the coal sulfur is converted to H_2S. Studies of the reaction $[S]_{org} + H_2 = H_2S$, have been published by Consol.(6,7)

In addition to being a CO_2 acceptor, CaO also is an efficient H_2S acceptor through the reaction,

$$CaO + H_2S = CaS + H_2O. \qquad (1)$$

Consol has patented processes (U.S. 2,824,047 and 3,117,918) for desulfurization of carbonaceous materials based on the above reaction. The acceptor also removes COS by the reaction,

$$CaO + COS = CaS + CO_2. \qquad (2)$$

Unpublished data by Consol show that a high degree of COS removal occurs by reaction (2).

For all three cases, equilibrium calculations show that about 96% of the coal sulfur is converted to CaS in the gasifier. On each cycle through the gasifier, about 2 mol % of the CaO in the recirculating acceptor stream is converted to CaS in the two high pressure cases. In the low pressure case, the acceptor circulation rate is much greater and conversion to CaS is about 0.7%.

A compilation of thermodynamic data for pertinent sulfur reactions and the procedure for equilibrium calculations are given in Appendix C.

The first step in recovery of the coal sulfur is regeneration of the CaS formed in the gasifier. Regeneration, and subsequent recovery of the sulfur, are described individually below for each of the three cases.

Sulfur Recovery

Probably, the best long-term method for disposal of recovered coal sulfur is conversion to elemental sulfur. The solid sulfur could be sold or, in the case of poor market conditions, could be stockpiled cheaply without causing pollution.

In general, technology of the processes to be described are not fully developed, for the conditions used. Design basis and assumptions are given for the individual cases.

Case I

Sulfur Rejection from CaS in the $CaCO_3$ Regenerator

In the regenerator the CaS content of the incoming acceptor is oxidized to $CaSO_4$ by both air and CO_2,

$$CaS \xrightarrow[CO_2]{O_2} CaSO_4$$

Then the following sequence takes place: The $CaSO_4$ decomposes by the reaction, $CaSO_4 = CaO + SO_3$. The SO_3 is unstable and decomposes by the reaction, $SO_3 = SO_2 + 1/2\ O_2$. The O_2 then reacts with unconverted CaS by the reaction, $1/4\ CaS + 1/2\ O_2 = 1/4\ CaSO_4$. The sum of these reactions is

$$3/4\ CaSO_4 + 1/4\ CaS = CaO + SO_2 \qquad (3)$$

At Case I conditions, the mol ratio, incoming CaS/exit gas, is such that the equilibrium partial pressure of SO_2 in reaction (3) would be reached after only about one-third of the CaS had reacted. To allow completion of the reaction, the regenerator is operated with a slight deficiency of combustion air to provide CO in the upper part of the fluidized bed which then reacts with SO_2 via

$$2\ CO + SO_2 = 1/2\ S_2 + 2\ CO_2, \qquad (4)$$

thereby decreasing the SO_2 partial pressure to about 70% of the equilibrium value in reaction (3). Details of the calculations which determine the necessary CO partial pressure are given in Appendix C.

The technical basis for this method of sulfur rejection was established during our studies for the OCR,(1) and during some preliminary work in which the CO_2 Acceptor Gasification Process was operated as a "CO-maker".(4) However, in those studies in which low-sulfur, low-rank Western coals were used, the partial pressures of SO_2 and S_2 in the regenerator were considerably lower than those required for Case I. The possibility exists that sulfur vapor will react with residual $CaSO_4$ in the acceptor or with the regenerated CaO via,

$$S_2 + CaSO_4 = CaS + 2\ SO_2, \qquad (5)$$

$$3/4\ S_2 + CaO = CaS + 1/2\ SO_2, \qquad (6)$$

with the result that the necessary rejection of sulfur from CaS will not occur. The kinetics of reactions (5) and (6) must be slow relative to the other sulfur reactions for successful use of the method. Study of the foregoing reactions at Case I conditions will be a major part of Phase II-B of the CAP contract.

Recovery of Elemental Sulfur

A flow diagram of the sulfur recovery section is shown in Figure 6. Mass and heat balances for all three cases are given in Section III-B-3. The regenerator exit gas, after removal of most of the coal ash and acceptor fines by cyclones, passes through heat exchangers which reduce the temperature to 472°F. Final clean up of particulates is by an electrostatic precipitator. About 34% of the total gas, after heat exchange, is passed over an alumina catalyst at 800°F in the reductor where the following reactions occur:

$$CO + H_2O = CO_2 + H_2$$

$$H_2 + 1/2\ S_2 = H_2S$$

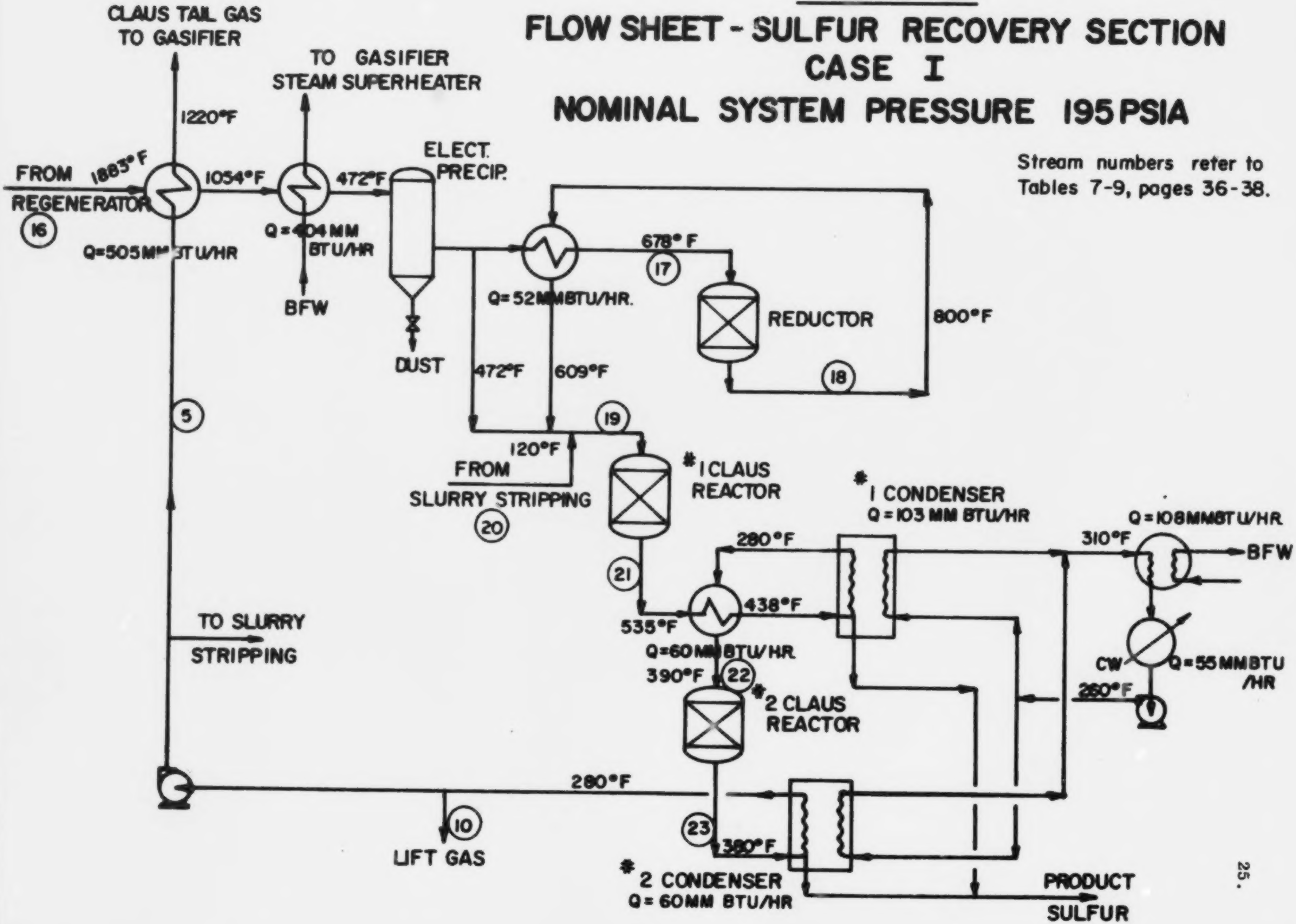
$$CO + 1/2\ S_2 = COS$$

$$2\ CO + SO_2 = 1/2\ S_2 + 2\ CO_2$$

FIGURE 6

FLOW SHEET - SULFUR RECOVERY SECTION
CASE I
NOMINAL SYSTEM PRESSURE 195 PSIA

Stream numbers refer to
Tables 7-9, pages 36-38.


CLAUS TAIL GAS TO GASIFIER
1220°F
FROM REGENERATOR
1883°F
⑯
Q=505 MMBTU/HR
1054°F
TO GASIFIER STEAM SUPERHEATER
Q=404 MM BTU/HR
BFW
472°F
ELECT. PRECIP.
DUST
472°F
Q=52 MMBTU/HR.
609°F
678° F
⑰
REDUCTOR
⑱
800°F
⑤
⑲
120°F
FROM SLURRY STRIPPING
⑳
# 1 CLAUS REACTOR
㉑
535°F
438°F
Q=60 MMBTU/HR.
390°F
㉒
# 2 CLAUS REACTOR
# 1 CONDENSER
Q = 103 MM BTU/HR
280°F
310°F
Q = 108 MMBTU/HR
BFW
CW
Q = 55 MMBTU /HR
260°F
TO SLURRY STRIPPING
280°F
⑩
LIFT GAS
㉓
380°F
# 2 CONDENSER
Q = 60 MM BTU/HR
PRODUCT SULFUR

After heat exchange, the reductor gas is combined with the remainder of the clean regenerator exit gas.

Discarded acceptor and overhead fines from the gasifier and regenerator contain CaS. These materials cannot be dumped because the CaS would hydrolyze on weathering, with release of H_2S to the atmosphere. These materials in all three cases, are slurried with water and contacted with a CO_2-rich gas in a stripping column where the following reaction occurs:

$$CaS + CO_2 + H_2O_{(1)} = CaCO_3 + H_2S.$$

The stripper gas then is combined with the regenerator gas.

The mol ratio, $(H_2S + COS)/SO_2$ in the combined gas streams now is 2.0, the optimum ratio for the Claus reactors which produce elemental sulfur by the reactions,

$$2\ H_2S + SO_2 = 3\ S + 2\ H_2O$$

$$2\ COS + SO_2 = 3\ S + 2\ CO_2$$

The tail gas from the No. 2 sulfur condenser is recompressed, reheated to 1220°F, and sent to the bottom of the gasifier.

Very high conversions to elemental sulfur are not of paramount importance in this method for sulfur recovery. Residual SO_2, H_2S, and COS in the Claus tail gas would be converted to CaS in the gasifier with the result that slightly more CO would be required in the regenerator to reject the incremental sulfur in the acceptor. In the heat and material balance calculations, the sulfur compounds in the Claus tail gas were ignored.

Unpublished work by Consol has shown that the reductor reactions very nearly approach equilibrium over an active alumina catalyst at 800°F at a space velocity of 4000 SCFH gas/CF catalyst bed. This work, done at atmospheric pressure, also showed that reactions involving CO and H_2 become very slow at temperatures below about 600°F. For calculations at Claus reactor temperatures, CO was assumed to behave as an inert gas. Claus reactor technology, also at atmospheric pressure, is well established. Since the mechanisms of the reductor and Claus reactions are not known, the effect of operation at elevated pressures could not be determined. Therefore, the space velocities used in Case I were those based on atmospheric pressure data.

Details of the reductor and Claus reactor calculations are given in Appendix C.

Case II

Sulfur Rejection from CaS in an External Reactor

Although Case I embodies the preferred method of sulfur rejection from the standpoint of cost and simplicity, an alternate method was desired in the event that forthcoming experimental studies show that the back reaction of sulfur vapor is an insurmountable problem.

In Case II, a very slight excess of air is used in the regenerator to prevent any substantial rejection of sulfur to the regenerator offgas. Most of the acceptor CaS is oxidized to $CaSO_4$. The regenerator gas is recompressed, without cooling, by a steam ejector and enters the bottom of the gasifier as shown in Figure 4. The sulfur is rejected from the acceptor by withdrawing one-third of the recirculating acceptor stream from the gasifier into an external fluidized bed reactor where the following reaction occurs:(18)

$$CaS + H_2O + CO_2 = CaCO_3 + H_2S. \qquad (7)$$

This acceptor stream, now free of sulfur, is then sent to the regenerator. Thus, the recirculating acceptor stream contains a sulfur "flywheel" in which the acceptor leaving the regenerator contains about 4 mol % of the total calcium in the form of $CaSO_4$ and the acceptor leaving the gasifier contains about 6 mol % of the total calcium in the form of CaS. In the gasifier the $CaSO_4$ in the recirculating acceptor is reduced to CaS.

A flow diagram for the Case II process is shown in Figure 7. In the external reactor, the acceptor is fluidized by a recirculating stream of steam and CO_2 in equimolar proportion at 1300°F. The ratio of inlet gas to acceptor is controlled to give a 90% approach to equilibrium in reaction (7) in the exit gas. Nominal retention time is one hour in the fluidized bed. One mol of CO_2 for each mol of CaS is supplied from a hot potassium carbonate unit which processes about 6.5% of the regenerator offgas, as shown in Figure 4.

The reactor offgas, after removal of most of the acceptor fines by cyclones, passes through heat exchangers which reduce the temperature to 480°F. Final cleanup is by an electrostatic precipitator.

Kinetic and equilibrium restrictions in reaction (7) limit the H_2S content in the reactor offgas to about 3.7 mol %. Lower reactor temperatures would give a higher concentration of H_2S, but unpublished work by Consol has shown that the kinetics of reaction (7) are poor at temperatures below about 1200°F.

The low concentration of H_2S in the reactor offgas, coupled with the high concentration of steam, provides an undesirable feedstock for sulfur recovery via the Claus process. For Case II another method of sulfur recovery was chosen.

Recovery of Elemental Sulfur

The reactor offgas is passed concurrently with a stream of aqueous H_2SO_3 through a packed tower where the Wackenroder reaction(5) takes place:

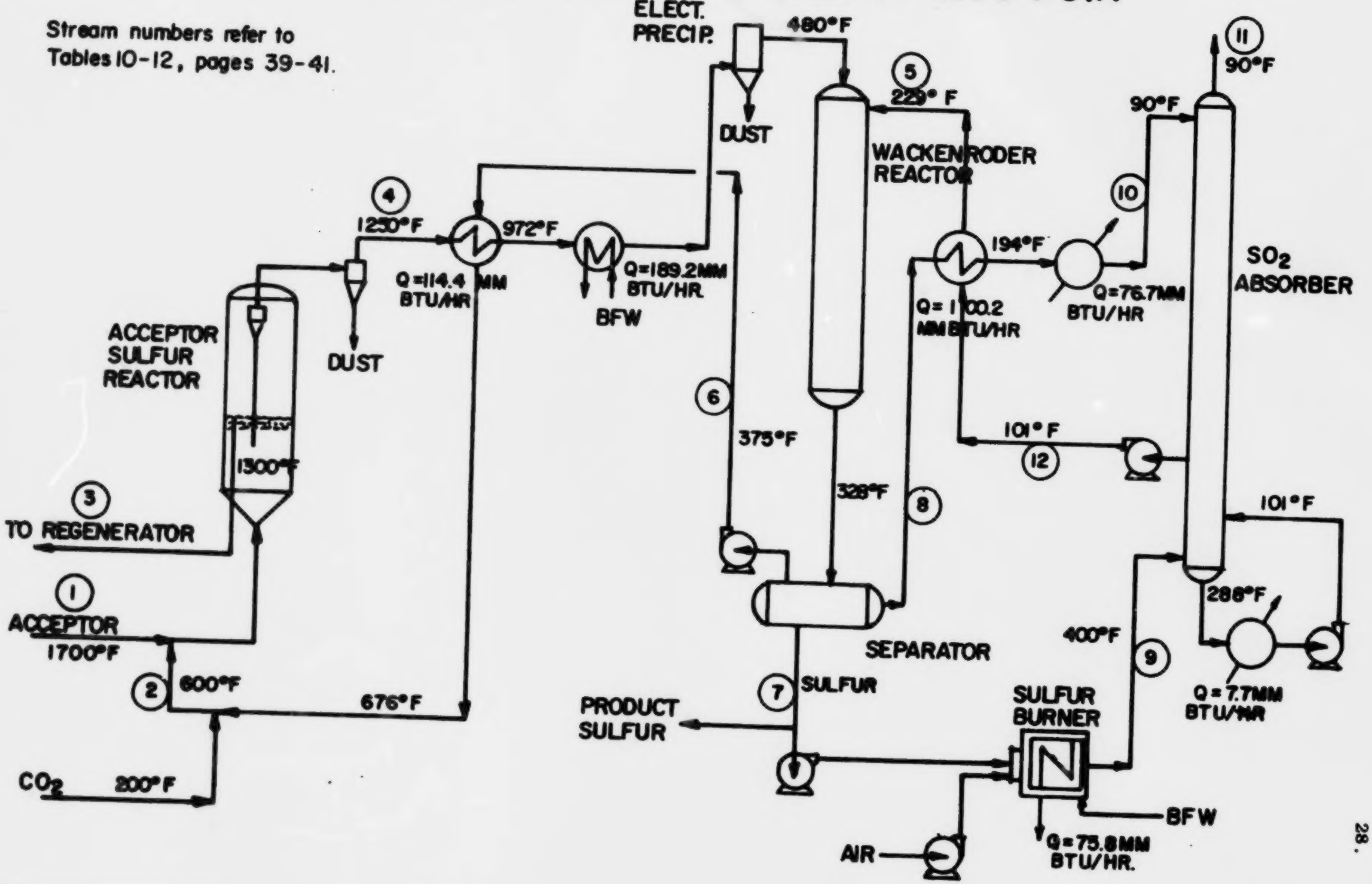
$$2\ H_2S + H_2SO_3 = 3\ S_{(1)} + 3\ H_2O. \qquad (8)$$

To avoid the very large heat duty involved in the condensation and re-evaporation of steam in the recirculating steam-CO_2 stream, reaction (8) is carried out at the saturation pressure. The reaction products pass to a separator where liquid water and liquid sulfur are removed. The steam-CO_2 stream is recompressed and sent back to the acceptor sulfur reactor. The water stream, after cooling, is sent to a tower where aqueous H_2SO_3 is formed by absorption of SO_2 produced by combustion of about one-third of the liquid sulfur produced from the Wackenroder reactor.

FIGURE 7

FLOW SHEET - SULFUR REMOVAL & RECOVERY SECTION
CASE II
NOMINAL SYSTEM PRESSURE 206 PSIA

Stream numbers refer to
Tables 10-12, pages 39-41.


ELECT.
PRECIP.
480°F
⑤
229° F
⑪
90°F
90°F
DUST
WACKENRODER
REACTOR
④
1250°F
972°F
⑩
Q=114.4 MM
BTU/HR
Q=189.2MM
BTU/HR
194°F
SO2
ABSORBER
BFW
Q=76.7MM
BTU/HR
ACCEPTOR
SULFUR
REACTOR
Q= 100.2
MM BTU/HR
DUST
⑥
375°F
1300°F
101° F
⑫
③
TO REGENERATOR
328°F
⑧
101°F
①
ACCEPTOR
1700°F
288°F
400°F
SEPARATOR
②
600°F
676°F
⑦
SULFUR
⑨
SULFUR
BURNER
Q = 7.7MM
BTU/HR
PRODUCT
SULFUR
CO2
200°F
BFW
AIR
Q=75.8MM
BTU/HR.

The kinetics of reaction (8) are not well established at the conditions used. However, as in Case I, very high conversion to liquid sulfur is not of paramount importance. By operating with a deficiency of H_2SO_3, the unconverted sulfur in the form of H_2S would be recirculated with the steam-CO_2 stream to the acceptor sulfur reactor which then would require a higher ratio of $(H_2O + CO_2)/CaS$ than is shown in the mass and heat balances.

Case III (Low Pressure Case)

Sulfur Rejection from CaS

At the low system pressure in Case III, the recirculating acceptor stream serves only as an H_2S acceptor and heat carrier, i.e., no $CaCO_3$ is formed. The regenerator is operated with stoichiometric air and sulfur is rejected by the reaction sequence:

$$\begin{array}{rcl} 3/4\ CaS + 3/2\ O_2 & = & 3/4\ CaSO_4 \\ \underline{3/4\ CaSO_4 + 1/4\ CaS} & \underline{=} & \underline{CaO + SO_2} \\ CaS + 3/2\ O_2 & = & CaO + SO_2 \end{array} \qquad (3)$$

as in Case I. However, no equilibrium restriction on SO_2 partial pressure exists because of the low system pressure.

Recovery of Elemental Sulfur

A flow diagram for Case III is shown in Figure 8. Regenerator offgas is cleaned and cooled as in Case I. The regenerator gas is mixed with 3.35% of the total product fuel gas and is passed over an active alumina catalyst at 800°F in the reductor. This amount of fuel gas provides sufficient CO and H_2 to give a mol ratio of $(H_2S + COS)/SO_2 = 2.0$ in the reductor product gas.

Elemental sulfur is formed in a single-stage Claus reactor. Calculations showed that a second Claus stage would give only a small increment in overall sulfur recovery.

The Claus tail gas must be sent to the boiler for incineration of the residual sulfur compounds. Because the Claus tail gas is not recycled to the process as in Case I, the overall sulfur clean-up level drops from 96% to 86%.

3. Material and Heat Balances - Gasification Section

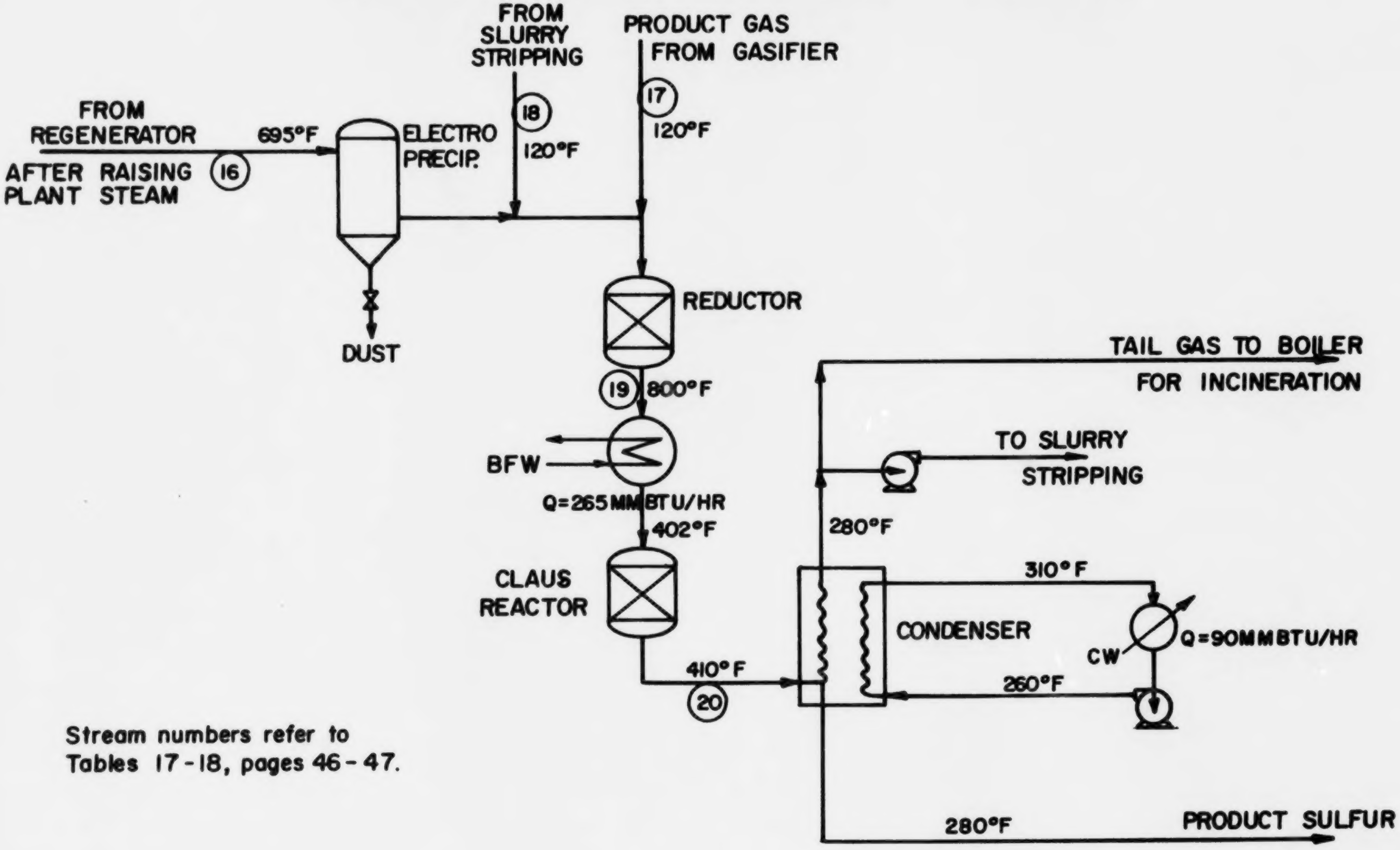
Basic Pressure Case I

Complete mass, elemental, and heat balances have been computed for all the major processing steps incorporated into the basis pressure case flow sheet as described previously. The overall mass and heat balance is summarized in Table III. The cold gas efficiency is:

$$\frac{10,286.6\ \text{MM Btu/hr}}{13,812.4\ \text{MM Btu/hr}}\ (100) = 74.5\%$$

FIGURE 8

FLOW DIAGRAM - SULFUR RECOVERY SECTION
CASE III
NOMINAL SYSTEM PRESSURE 20 PSIA


FROM
SLURRY
STRIPPING
PRODUCT GAS
FROM GASIFIER
FROM
REGENERATOR
AFTER RAISING
PLANT STEAM
16
695°F
ELECTRO
PRECIP.
18
120°F
17
120°F
DUST
REDUCTOR
19
800°F
BFW
Q=265 MMBTU/HR
402°F
CLAUS
REACTOR
410°F
20
TAIL GAS TO BOILER
FOR INCINERATION
TO SLURRY
STRIPPING
280°F
310°F
CONDENSER
CW
Q=90 MMBTU/HR
260°F
280°F
PRODUCT SULFUR

Stream numbers refer to
Tables 17-18, pages 46-47.

TABLE 3

Overall Mass and Heat Balance
Gasification Section-Case I

Basis: 1 Hour Datum: 60°F $H_2O_{(1)}$

	Pounds	Mols	Mol %	Temp. °F	ΔH MM Btu	ΔHc HHV MM Btu	ΔH + ΔHc MM Btu
Input							
Feed coal (6% moisture)	1,157,000	--	--	60	x	13,812.4	13,812.4
Air	3,949,200	137,095	--	398	335.8	x	335.8
Boiler feed water	731,300	--	--	60	x	x	335.8
Make-up Acceptor	77,300	--	--	60	x	x	x
Totals	5,914,800				335.8	13,812.4	14,148.2
Output							
Product Gas							
CH_4	48,400	3,017	1.3	1300	48.1	1,154.8	1,202.9
H_2	71,800	35,624	15.2	1300	310.7	4,388.0	4,698.7
CO	1,069,500	38,182	16.2	1300	348.6	4,646.8	4,995.4
CO_2	932,700	21,194	9.0	1300	296.7	x	296.7
N_2	3,026,900	108,041	46.0	1300	976.5	x	976.5
NH_3	8,700	509	0.2	1300	6.9	x	976.5
H_2S	1,800	54	x	1300	0.6	83.9	90.8
H_2O (v)	510,900	28,359	12.1	1300	849.7	13.1	13.7
Sub-total	5,670,700	234,980	100.0		2,837.8	x	849.7
						10,286.6	13,124.4
Unburned Char and Ash	152,800	--	--	120	1.1	271.7	272.8
Product $Sulfur_{(liq)}$	44,800	--	--	280	2.8	178.9	181.7
Spent Acceptor	45,900	--	--	120	0.1	x	0.1
Net Heat of Reaction and Sensible Heat Losses							569.2
Totals	5,914,200	--	--		2,841.8	10,737.2	14,148.2

A more realistic evaluation of the thermal efficiency of the process is obtained by relating the total heat content of the product gas and of the product sulfur to the total heat input:

$$\frac{13,306.1 \text{ MM Btu/hr}}{14,148.2 \text{ MM Btu/hr}} (100) = 94.0\%$$

Detailed mass, elemental, and heat balances for the following operations are shown as follows:

Preoxidation - Table 4
Gasification - Table 5
Regeneration - Table 6
Reductor - Table 7
First-Stage Claus - Table 8
Second-Stage Claus - Table 9

Alternate Pressure Case II

As mentioned previously, the main difference between Cases I and II is the method by which the sulfur is released from the acceptor. Since the heat and material balance calculations showed that the energy contained by the product gas in Case II was within 1% of that in Case I, only the sulfur removal and recovery details are presented here. Detailed mass and heat balances for the novel acceptor processing cycle were then computed, and are shown as follows:

Acceptor Sulfur Reactor - Table 10
"Wackenroder" Reactor - Table 11
SO_2 Absorption Tower - Table 12

Atmospheric Pressure Case III

Again, complete mass, elemental, and heat balances have been computed for all the major processing steps incorporated into the basic flow sheet as described previously. The overall mass and heat balance is summarized in Table 13.

For this case, the cold gas efficiency is:

$$\frac{10,078.5 \text{ MM Btu/hr}}{13,812.4 \text{ MM Btu/hr}} (100) = 73.0\%$$

The thermal efficiency is expressed by the ratio of the total heat content of the product gas and of the product sulfur to the total heat input is:

$$\frac{11,627.4 \text{ MM Btu/hr}}{14,082.6 \text{ MM Btu/hr}} (100) = 82.5\%$$

Detailed mass, elemental, and heat balances for the following operations are shown as follows:

Preoxidation - Table 14
Gasification - Table 15
Regeneration - Table 16
Reductor - Table 17
Claus Stage - Table 18

TABLE 4

Mass & Heat Balance
Preoxidizer
Case I

- See Figure 3

Basis: 1 Hour Datum: 60°F. $H_2O_{(l)}$

	Pounds	Mols	Mol %	Elemental Balance, Pounds						Temp. °F	ΔH or Cp	ΔH MM Btu	Heat of Combustion HHV MM Btu
				H	C	N	O	S	Ash				
Input													
① **Feed Coal**													
MF Coal	1,087,600	-	-	52,200	759,200	13,000	82,600	46,800	133,800	60	-	x	13,812.4
Water(l)	69,400	-	-	7,700	x	x	61,700	x	x	"	-	x	x
Sub-Total	1,157,000	-	-	59,900	759,200	13,000	144,300	46,800	133,800			x	13,812.4
② **Air**													
O_2	119,800	3,744	20.9	x	x	x	119,800	x	x	398	(2365 Btu/mol (	42.2	x
N_2	394,700	14,087	78.7	x	x	394,700	x	x	x	"	(" (		x
H_2O (v)	1,400	77	0.4	200	x	x	1,200	x	x	"	1210.8 Btu/lb	1.7	x
Sub-Total	515,900	17,908	100.0	200	x	394,700	121,000	x	x			43.9	x
Heat Release by Oxidation (2921 mols O_2 x 185,000 Btu/mol)												540.4	
Totals	1,672,900	-	-	60,100	759,200	407,700	265,300	46,800	133,800			584.3	
Output													
③ **Preoxidized Coal**	1,009,000	-	-	44,100	705,200	12,400	68,100	45,100	133,800	800	266 Btu/lb	268.4	12,653.9
③ **Preoxidizer Vapor**													
Tar + Light Oil (v)	28,300	189	0.8	2,100	23,300	300	2,100	500	x	800	533 Btu/lb	15.1	452.4
CH_4	1,600	98	0.4	400	1,200	x	x	x	x	"	8,192 Btu/mol	0.8	37.5
C_2H_6	800	26	0.1	200	600	x	x	x	x	"	13,569 "	0.4	17.6
H_2	900	461	1.9	900	x	x	x	x	x	"	5,168 "	2.4	56.8
CO	10,200	364	1.5	x	4,400	x	5,800	x	x	"	5,288 "	1.9	44.3
CO_2	90,000	2,046	8.1	x	24,500	x	65,500	x	x	"	7,784 "	15.9	x
O_2	26,400	824	3.4	x	x	x	26,400	x	x	"	5,491 "	4.5	x
N_2	395,000	14,099	58.2	x	x	395,000	x	x	x	"	5,251 "	74.0	x
H_2S	1,300	39	0.2	100	x	x	x	1,200	x	"	6,525 "	0.3	9.5
H_2O (v)	109,400	6,068	25.1	12,300	x	x	97,100	x	x	"	1402.5 Btu/lb	153.4	x
Sub Total	663,900	24,214	100.0	16,000	54,000	395,300	196,900	1,700	x			268.7	618.1
Heat Loss												47.2	
Totals	1,672,900	-	-	60,100	759,200	407,700	265,300	46,800	133,800			584.3	

TABLE 5

Mass & Heat Balance
Gasifier
Case I

- See Figure 3

Basis: 1 Hour Datum: 60°F $H_2O_{(l)}$

	Pounds	Mols	Mol %	Elemental Balance, Pounds							Temp. °F	ΔH or Cp	ΔH MM Btu	Heat of Combustion HHV MM Btu
				H	C	N	O	S	Ash	MgO·Ca				
Input														
③ **Preoxidized Coal**	1,009,000	-	-	44,100	705,200	12,400	68,400	45,100	133,800	x	800	266 Btu/lb	268.4	12,653.9
③ **Preoxidizer Vapor**	683,900	24,214	-	16,000	54,000	395,300	196,900	1,700	x	x	800	(Table 4)	268.7	618.1
⑪ **Air**														
O_2	402,600	12,581	20.9	x	x	x	402,600	x	x	x	398	(2365 Btu/mol	(141.7	x
N_2	1,326,600	47,351	78.7	x	x	1,326,600	x	x	x	x	"	("	(	x
H_2O (v)	4,700	259	0.4	500	x	x	4,200	x	x	x	"	1210.8 Btu/lb	5.7	x
Sub-Total	1,733,900	60,191	100.0	500	x	1,326,600	406,800	x	x	x			147.4	x
④ **Steam**	731,300	40,595	-	81,800	x	x	649,500	x	x	x	1200	1607.8 Btu/lb	1,175.9	x
⑤ **Claus Tail Gas**														
CO	32,800	1,171	1.7	x	14,100	x	18,700	x	x	x	1200	8,344 Btu/mol	9.8	142.5
CO_2	931,600	21,167	30.4	x	254,200	x	677,400	x	x	x	"	12,715 "	269.1	x
H_2	100	24	x	100	x	x	x	x	x	x	"	8,004 "	0.2	3.0
N_2	1,300,900	46,433	66.8	x	x	1,300,900	x	x	x	x	"	8,263 "	383.7	x
SO_2	300	5	x	x	x	x	100	200	x	x	"	13,107 "	0.1	x
H_2S	300	10	x	x	x	x	x	300	x	x	"	10,574 "	0.1	3.4
H_2O (v)	13,900	771	1.1	1,500	x	x	12,400	x	x	x	"	1607.8 Btu/lb	22.3	x
Sub-Total	2,279,900	69,581	100.0	1,600	268,300	1,300,900	708,600	500	x	x			685.3	147.9
⑥ **Acceptor**														
MgO·CaO	7,399,800	76,761	-	x	x	x	1,228,200	x	x	6,171,600	1883	23.2 Btu/mol-°F	3,246.5	x
Inert	1,305,800	-	-	x	x	x	x	x	1,305,800	x	"	0.255 Btu/lb-°F	607.0	x
Sub-Total	8,705,600	-	-	x	x	x	1,228,200	x	1,305,800	6,171,600			3,853.5	x
Heat of Reaction														
MgO·CaO to MgO·$CaCO_3$ (9,699 mols x 76,200 Btu/mol)													739.1	
MgO·CaO to MgO·CaS (1,416 mols x 45,500 Btu/mol)													64.4	
Heat of Combustion (Reactants - Products ex Acceptor)													738.3	
Totals	15,123,600	-	-	114,000	1,027,500	3,035,200	3,258,400	47,300	1,439,600	6,171,600			7,941.0	
Output														
⑦ **Char**	293,700	-	-	1,200	159,200	x	x	x	133,300	x	1700	545 Btu/lb	160.1	2,349.9
⑧ **Char Loss**	3,600	-	-	100	2,500	x	300	200	500	x	1700	545 Btu/lb	1.9	45.1
⑫ **Acceptor**														
MgO·$CaCO_3$	1,361,800	9,699	-	x	116,500	x	465,500	x	x	779,800	1700	38.7 Btu/mol-°F	815.6	x
MgO·CaS	159,200	1,416	-	x	x	x	x	45,400	x	113,800	"	23.0 "	53.4	x
MgO·CaO	6,328,300	65,646	-	x	x	x	1,050,300	x	x	5,278,000	"	23.0 "	2,476.1	x
Inert	1,305,800	-	-	x	x	x	x	x	1,305,800	x	"	0.251 Btu/lb-°F	537.5	x
Sub-Total	9,155,100	-	-	x	116,500	x	1,515,800	45,400	1,305,800	6,171,600			3,882.6	x
⑨ **Gas**														
CH_4	48,400	3,017	1.3	12,200	36,200	x	x	x	x	x	1700	23,134 Btu/mol	69.8	1,154.8
H_2	71,800	35,624	15.2	71,800	x	x	x	x	x	x	"	11,641 "	414.7	4,388.0
CO	1,069,500	38,182	16.2	x	458,600	x	610,900	x	x	x	"	12,347 "	471.4	4,646.8
CO_2	932,700	21,194	9.0	x	254,500	x	678,200	x	x	x	"	19,290 "	408.8	x
N_2	3,026,900	108,041	46.0	x	x	3,026,900	x	x	x	x	"	12,213 "	1,319.5	x
NH_3	8,700	509	0.2	1,500	x	7,200	x	x	x	x	"	19,203 "	9.8	83.9
H_2S	1,800	54	x	100	x	x	x	1,700	x	x	"	16,094 "	0.9	13.1
H_2O(v)	510,900	28,339	12.1	57,200	x	x	453,700	x	x	x	"	1884.9 Btu/lb	963.0	x
Sub-Total	5,670,700	234,980	100.0	142,800	749,300	3,034,100	1,742,800	1,700	x	x			3,657.9	10,286.6
Heat of Reaction														
Heat of Combustion (1,416 mols H_2S to CaS)													342.6	
Heat Loss													95.9	
Totals	15,123,100	-	-	144,100	1,027,500	3,034,100	3,258,900	47,300	1,439,600	6,171,600			7,941.0	

TABLE 6

Mass & Heat Balance
Regenerator
Case I

- See Figure 3

Basis: 1 Hour   Datum: 60°F. $H_2O_{(1)}$

	Pounds	Mols	Mol %	Elemental Balance, Pounds							Temp. °F	ΔH or Cp	ΔH MM Btu	Heat of Combustion HHV MM Btu
				H	C	N	O	S	Ash	MgO·Ca				
Input														
⑦ Char	293,700	-	-	1,200	159,200	x	x	x	133,300	x	1700	545 Btu/lb	160.1	2,349.9
⑩ Lift Gas														
CO	2,100	76	1.7	x	900	x	1,200	x	x	x	280	1537 Btu/mol	0.1	9.2
CO_2	60,700	1,374	30.4	x	16,600	x	44,100	x	x	x	"	2072 "	2.8	x
H_2	x	1	x	x	x	x	x	x	x	x	"	1528 "	x	0.1
N_2	84,700	3,022	66.8	x	x	84,700	x	x	x	x	"	1534 "	4.6	x
H_2O (v)	900	51	1.1	100	x	x	800	x	x	x	"	1158.4 Btu/lb	1.0	x
Sub-Total	148,100	4,528	100.0	100	17,500	84,700	46,100	x	x	x			8.5	9.3
⑭ Air														
O_2	394,700	12,336	20.9	x	x	x	394,700	x	x	x	398	(2365	(138.9	x
N_2	1,300,100	46,406	78.7	x	x	1,300,100	x	x	x	x	"	(	(	x
H_2O (v)	4,600	254	0.4	500	x	x	4,100	x	x	x	"	1210.8 Btu/lb	5.6	x
Sub-Total	1,699,400	58,996	100.0	500	x	1,300,100	398,800	x	x	x			144.5	x
⑫ Acceptor														
$MgO \cdot CaCO_3$	1,354,900	9,650	-	x	115,900	x	463,200	x	x	775,800	1700	38.7 Btu/mol-°F	612.5	x
$MgO \cdot CaS$	158,600	1,410	-	x	x	x	x	45,200	x	113,400	"	23.0 "	53.2	x
$MgO \cdot CaO$	6,296,600	65,317	-	x	x	x	1,045,100	x	x	5,251,500	"	23.0 "	2,463.7	x
Inert	1,299,300	-	-	x	x	x	x	x	1,299,300	x	"	0.251 Btu/lb-°F	534.8	x
Sub-Total	9,109,400	-	-	x	115,900	x	1,508,300	45,200	1,299,300	6,140,700			3,664.2	x
⑬ Make-up Acceptor														
$MgCO_3 \cdot CaCO_3$	70,800	384	-	x	9,200	x	30,700	x	x	30,900	60	--	x	x
Inert	6,500	-	-	x	x	x	x	x	6,500	x	"	--	x	x
Sub-Total	77,300	-	-	x	9,200	x	30,700	x	6,500	30,900			x	x
Heat of Reaction														
$MgO \cdot CaS$ to $MgO \cdot CaO$ (1410 mols x 196,600 Btu/mol)													277.2	
Heat of Combustion (Reactants - Products ex Acceptor)													1,731.4	
Totals	11,328,200	-	-	1,800	301,800	1,384,800	1,983,900	45,200	1,439,100	6,171,600			5,985.9	
Output														
⑮ Char	149,200	-	-	x	15,900	x	x	x	133,300	x	1883	490 Btu/lb	73.1	226.6
⑥ Acceptor														
$MgO \cdot CaO$	7,399,800	76,761	-	x	x	x	1,228,200	x	x	6,171,600	1883	23.2 Btu/mol-°F	3,246.5	x
Inert	1,305,800	-	-	x	x	x	x	x	1,305,800	x	"	0.255 Btu/lb-°F	607.0	x
Sub-Total	8,705,600	-	-	x	x	x	1,228,200	x	1,305,800	6,171,600			3,853.5	x
⑯ Gas														
CO	52,300	1,867	2.5	x	22,400	x	29,900	x	x	x	1883	13,852 Btu/mol	25.9	227.2
CO_2	964,900	21,925	29.2	x	263,300	x	701,600	x	x	x	"	21,780 "	477.5	x
H_2	100	38	0.1	100	x	x	x	x	x	x	"	13,007 "	0.5	4.7
N_2	1,385,500	49,455	66.0	x	x	1,385,500	x	x	x	x	"	13,700 "	677.5	x
SO_2	20,800	325	0.4	x	x	x	10,400	10,400	x	x	"	22,021 "	7.1	x
S_2	34,400	536	0.7	x	x	x	x	34,400	x	x	"	15,684 "	8.4	166.7
H_2S	200	5	x	x	x	x	x	200	x	x	"	18,215 "	0.1	1.2
COS	300	6	x	x	100	x	100	100	x	x	"	23,051 "	0.1	1.4
H_2O (v)	11,700	811	1.1	1,600	x	x	13,100	x	x	x		1987.4 Btu/lb	29.1	x
Sub-Total	2,473,200	74,971	100.0	1,700	285,800	1,385,500	755,100	45,100	x	x			1,226.2	401.2
Heat of Reaction														
$MgO \cdot CaCO_3$ to $MgO \cdot CaO$ (9,650 mols x 76,200 Btu/mol)													735.3	
$MgCO_3 \cdot CaCO_3$ to $MgO \cdot CaO$ (384 mols x 127,400 Btu/mol)													48.9	
Heat Loss													48.9	
Totals	11,328,000	-	-	1,700	301,700	1,385,500	1,983,300	45,100	1,439,100	6,171,600			5,985.9	

TABLE 7

Mass & Heat Balance
Reductor
Case I - See Figures 3 and 6

Basis: 1 hour Datum: 60°F $H_2O_{(l)}$

	Pounds	Mols	Temp °F	ΔH or Cp	ΔH MM Btu
Input					
⑰ Regenerator Gas					
CO	17,530	625.8	678	4385 Btu/mol	2.74
CO_2	323,400	7349.3		6356 "	46.71
H_2	25	12.7		4320 "	.05
N_2	464,500	16,577		4366 "	72.38
SO_2	6,976	108.9		6668 "	.73
S_2	11,370	177.3		5113 "	.91
H_2S	61	1.8		5415 "	.01
COS	132	2.2		6996 "	.02
H_2O (v)	4,916	272.8		1344.3 Btu/lb.	6.61
Sub-total	828,910	25,128			130.16
Heat of Reaction: Heat of Combustion (Reactants - Products)					24.97
Totals	828,910	25,128			155.13
Output					
⑱ Reductor Gas					
CO	148	5.3	800	5288 Btu/mol	.03
CO_2	343,100	7,795.9		7784 "	60.68
H_2	1	0.3		5168 "	x
N_2	464,500	16,577		5251 "	87.04
SO_2	2	0.03		8123 "	x
* S_2	1,551	24.19		6144 "	0.15
H_2S	8,288	243.5		6525 "	1.59
COS	10,540	175.7		8510 "	1.50
H_2O (v)	780	43.6		1404.7 Btu/lb.	1.10
Sub-total	828,910	24,866			152.09
Heat Loss					3.04
Totals	828,910	24,866			155.13

* Shown as S_2. Actual composition is 3.875 atoms S/mol

TABLE 8

Mass & Heat Balance
First-Stage Claus Reactor
Case I - See Figures 3 and 6

Basis: 1 hour Datum: 60°F $H_2O_{(1)}$

	Pounds	Mols	Temp °F	ΔH or Cp	ΔH MM Btu
Input					
⑲ Feed Gas					
CO	34,910	1,246.5	535	3348 Btu/mol	4.17
CO_2	984,600	22,372		4753 "	106.33
H_2	52	25.6		3309 "	.08
N_2	1,385,700	49,455		3334 "	164.88
SO2	13,840	216.1		4996 "	1.08
* S_8	24,100	93.97		18,887 "	1.77
H_2S	8,420	247.1		4051 "	1.00
COS	10,810	180.0		5238 "	0.94
H_2O (v)	10,540	584.8		1276.8 Btu/lb.	13.45
Sub-total	2,472,970	74,421			293.70
⑳ Gas from Slurry Stripping					
CO	110	4	120	x	x
CO_2	2600	59		542 Btu/mol	.03
N_2	4000	143		417 "	.06
H_2S	200	6		x	x
H_2O (v)	40	2		x	x
Sub-total	6950	214			.09
Heat of Reaction					
$SO_2 + 2\ COS = 3/2\ S_2 + 2\ CO_2$ (89.40 mols SO_2 x 8510 Btu/mol)					.76
$SO_2 + 2\ H_2S = 3/2\ S_2 + 2\ H_2O_{(1)}$ (98.65 mols SO_2 x 17,780 Btu/mol)					1.75
$4\ S_2 = S_8$ (71.59 mols S_8 x 178,000 Btu/mol)					12.78
Totals	2,479,920	74,635			309.08
Output					
㉑ Product Gas					
CO	35,020	1,250.5	535	3348 Btu/mol	4.19
CO_2	995,100	22,610		4753 "	107.48
H_2	52	25.6		3309 "	.08
N_2	1,389,700	49,598		3334 "	165.36
SO_2	1,634	25.51		4996 "	.13
** S_8	42,460	165.56		18,887 "	3.13
H_2S	1,690	49.80		4051 "	.20
COS	72	1.20		5138 "	.01
H_2O (v)	14,190	788.3		1276.8 Btu/lb	18.14
Sub-total	2,479,920	74,514			298.72
Heat Loss					10.36
Totals	2,479,920	74,514			309.08

* Shown as S_8. Actual composition is approx. 7 atoms S/mol

** Shown as S_8. Actual composition is 7.294 atoms S/mol

TABLE 9

Mass & Heat Balance
Second-Stage Claus Reactor
Case I - See Figures 3 and 6

Basis: 1 hour Datum: 60°F $H_2O(l)$

	Pounds	Mols	Temp. °F	ΔH or Cp	ΔH MM Btu
Input					
㉒ **Feed Gas**					
CO	35,020	1,250.5	390	2312 Btu/mol	2.89
CO_2	995,100	22,610		3193 "	72.21
H_2	52	25.6		2295 "	.06
N_2	1,385,700	49,598		2306 "	114.37
SO_2	1,634	25.51		3372 "	.09
H_2S	1,690	49.80		2765 "	.14
COS	72	1.20		x	x
H_2O (v)	14,190	788.3		1208.8 Btu/lb.	17.17
Sub-total	2,433,460	74,349			206.93
Heat of Reaction					
$SO_2 + 2\ COS = 3/2\ S_2 + 2\ CO_2$ (0.57 mols SO_2 x 8510 Btu/mol)					.01
$SO_2 + 2\ H_2S = 3/2\ S_2 = 2\ H_2O(l)$ (19.97 mol SO_2 x 17,780 Btu/mol)					.36
$4\ S_2 = S_8$ (7.69 mols S_8 x 178,500 Btu/mol)					1.37
Totals					208.67
Output					
㉓ **Product Gas**					
CO	35,020	1,250.5	380	2242 Btu/mol	2.80
CO_2	995,100	22,611		3090 "	69.88
H_2	52	25.6		2225 "	.06
N_2	1,385,700	49,598		2236 "	110.90
SO_2	318	4.96		3306 "	.02
* S_8	1,973	7.69		12,508 "	.10
H_2S	336	9.86		2684 "	.03
COS	4	.06		x	x
H_2O (v)	14,930	828.4		1204.2 Btu/lb	17.98
Sub-total	2,433,430	74,336			201.77
Heat Loss					6.90
Totals	2,433,430	74,336			208.67

* Shown as S_8. Actual composition is 7.518 atoms S/mol

TABLE 10

Mass & Heat Balance
Acceptor Sulfur Reactor
Case II

- See Figures 4 & 7

Basis: 1 hour   Datum: 60°F $H_2O_{(l)}$

	Pounds	Mols	Temp °F	ΔH or Cp	ΔH MM Btu
Input					
① **Acceptor**					
$MgO \cdot CaCO_3$	370,550	2,639.	1700	38.7 Btu/mol-°F	167.4
$MgO.CaS$	154,700	1,375.5	"	23.0 "	51.9
$MgO \cdot CaO$	1,627,400	16,882.	"	23.0 "	636.8
Inert	355,500	-	"	0.251 Btu/lb -°F	146.4
Sub-total	2,508,150	-			1,002.5
② **Gas**					
CO_2	852,400	19,368	600	5473 Btu/mol	106.0
H_2O (v)	348,950	19,368	"	1299.8 Btu/lb.	453.6
Sub-total	1,201,350	38,736			559.6
Heat of Reaction $MgO \cdot CaS + CO_2 + H_2O(l) = MgO \cdot CaCO_3 + H_2S$ (1375.5 mols x 30,620 Btu/mol)					42.1
Totals	3,709,500	-			1,604.2
Output					
③ **Acceptor**					
$MgO \cdot CaCO_3$	563,700	4,014.5	1300	36.9 Btu/mol-°F	183.9
$MgO \cdot CaO$	1,627,400	16,882.	"	22.6 "	472.4
Inert	355,500	-	"	0.242 Btu/lb-°F	106.9
Sub-total	2,546,600	-			763.2
④ **Gas**					
CO_2	791,850	17,992.5	1300	13,999 Btu/mol	251.9
H_2S	46,900	1,375.5	"	11,644 "	16.0
H_2O (v)	324,150	17,992.5	"	1661.8 Btu/lb	538.7
Sub-total	1,162,900	37,360.5			806.6
Heat Loss					34.4
Totals	3,709,500	-			1,604.2

TABLE 11

Mass & Heat Balance
"Wackenroder" Reactor
Case II

- See Figures 4 & 7

Basis: 1 hour   Datum: 60°F $H_2O_{(l)}$

	Pounds	Mols	Temp. °F	ΔH or Cp	ΔH MM Btu
Input					
④ Gas					
CO_2	791,850	17,992.5	480	4152 Btu/mol	74.7
H_2S	46,900	1,375.5	"	3557 "	4.9
H_2O (v)	324,150	17,992.5	"	1240.9 Btu/lb	402.2
Sub-total	1,162,900	37,360.5			481.8
⑤ Liquid					
H_2SO_3	56,450	688	229	1.0 Btu/lb-°F	132.3
$H_2O_{(l)}$	726,450	40,324	"		
Sub-total	782,900	41,012			132.3
Heat of Reaction $2\ H_2S + H_2SO_3 = 3\ S_{(l)} + 3\ H_2O_{(l)}$ (1375.5 mols H_2S x 43,900 Btu/mol)					60.4
Totals	1,945,800	-			674.5
Output					
⑥ Gas					
CO_2	791,850	17,992.5	328	2556 Btu/mol	46.0
H_2O (v)	348,800	19,361	"	1159.1 Btu/lb	404.3
Sub-total	1,140,650	37,353.5			450.3
Liquid					
⑦ Sulfur	66,150	2,063	328	0.2 Btu/lb-°F	3.5
⑧ $H_2O_{(l)}$	739,000	41,019	"	270.5 Btu/lb	199.0
Sub-total	805,150	43,082			202.5
Heat Loss					21.7
Totals	1,945,800	-			674.5

TABLE 12

Mass & Heat Balance
SO_2 Absorption Tower
Case II

- See Figures 4 & 7

Basis: 1 hour Datum: 60°F $H_2O_{(l)}$

	Pounds	Mols	Temp. °F	ΔH or Cp	ΔH MM Btu
Input					
⑨ **Gas**					
SO_2	44,500	695	400	3480 Btu/mol	2.4
N_2	73,250	2,614.5	"	2376 "	6.2
H2O (v)	500	29	"	1213.5 Btu/lb	0.6
Sub-total	118,250	3,338.5			9.2
⑩ **Water** $_{(l)}$	738,500		90	29.93 Btu/lb	22.1
Heat of Reaction					
$SO_2 + H_2O_{(l)} = H_2SO_3$ (aq)					
(688 mols x 13,400 Btu/mol)					9.2
Totals	856,750	-			40.5
Output					
⑪ **Gas**					
SO_2	450	7	90	284 Btu/mol	x
N_2	73,250	2,614.5	"	208 "	0.55
H_2O (v)	150	8.5	"	1072.8 Btu/lb	0.15
Sub-total	73,850	2,630.			0.7
⑫ **Liquid**					
H_2SO_3	56,450	688	101	1.0 Btu/lb-°F	32.1
$H_2O_{(l)}$	726,450	40,324	"		
Sub-total	782,900	41,012			32.1
Heat Rejected to Cooling Water					7.7
Totals	856,750	-			40.5

TABLE 13

Overall Mass and Heat Balance Gasification Section-Case III

Basis: 1 hour Datum: 60°F $H_2O_{(1)}$ - See Figure 5

	Pounds	Mols	Mol %	Temp. °F	ΔH MM Btu	ΔHc HHV MM Btu	ΔH + ΔHc MM Btu
Input							
Feed coal (6% moisture)	1,157,000	--	--	60	--	13,812 4	13,812.4
Air	4,054,800	140,996	--	300-315	270.2	--	270.2
Boiler feed water	614,700	--	--	60	--	--	--
Make-up Acceptor	84,800	--	--	60	--	--	--
Process water	200	--	--	60	--	--	--
Totals	5,911,500	140,996	--		270.2	13,812.4	14,082.6
Output							
Product Gas							
CH_4	3,700	232	0.2	980	2.5	88.6	91.1
H_2	88,500	43,916	27.6	980	282.8	5,409.5	5,692.3
CO	1,034,300	36,928	23.3	980	245.4	4,494.3	4,739.7
CO_2	403,500	9,167	5.8	980	91.3	--	91.3
N_2	1,469,300	52,445	33.0	980	345.6	--	345.6
NH_3	8,100	479	0.3	980	4.6	78.9	83.5
H_2S	1,100	30	--	980	0.2	7.2	7.4
H_2O (v)	281,700	15,636	9.8	980	421.2	--	421.2
Sub-total	3,290,200	158,833	100.0		1,393.6	10,078.5	11,472.1
In-Process Fuel Gas	35,200	1,700	--		14.9	108.0	122.9
Unburned Char-Ash	164,500	--	--	120	1.2	440.6	441.8
Product Sulfur $_{(liq)}$	38,300	--	--	280	2.4	152.9	155.3
Claus Plant Tail Gas							
CH_4	100	7		280	--	2.7	2.7
CO_2	654,200	14,865		280	30.8	--	30.8
N_2	1,614,100	57,615		280	88.4	--	88.4
H_2S	3,300	97		280	0.2	23.4	23.6
SO_2	3,100	49		280	0.1	--	0.1
S_8	1,700	6		280	0.1	6.6	6.7
H_2O	57,800	3,213			67.0	--	67.0
Sub-total	2,334,300	75,852			186.6	32.7	219.3
Spent Acceptor	47,900	--	--	120	0.1	--	0.1
Net Heat of Reaction and Sensible Heat Losses							1,671.1
Totals	5,910,400				1,599.6	10,812.7	14,082.6

TABLE 14

Mass & Heat Balance
Preoxidizer
Case III

- See Figure 5

Basis: 1 hour Datum: 60°F $H_2O_{(l)}$

	Pounds	Mols	Mol %	Elemental Balance, Pounds						Temp. °F	ΔH or Cp	ΔH MM Btu	Heat of Combustion HHV MM Btu
				H	C	N	O	S	Ash				
Input													
① **Feed Coal**													
MF Coal	1,087,600	-	-	52,200	759,200	13,000	82,600	46,800	133,800	60	-	x	13,812.4
Water$_{(l)}$	69,400	-	-	7,700	x	x	61,700	x	x	"	-	x	x
Sub-total	1,157,000	-	-	59,900	759,200	13,000	144,300	46,800	133,800			x	13,812.4
② **Air**													
O_2	122,500	3,829	20.8	x	x	x	122,500	x	x	315	1816 Btu/mol	7.0	x
N_2	403,600	14,405	78.3	x	x	403,600	x	x	x	"	1779 "	25.6	x
H_2O (v)	2,900	161	0.9	300	x	x	2,600	x	x	"	1174.5 Btu/lb	3.4	x
Sub-total	529,000	18,395	100.0	300	x	403,600	125,100	x	x			36.0	x
Heat Release by Oxidation (2987 mols O_2 x 185,000 Btu/mol)												552.6	
Totals	1,686,000	-	-	60,200	759,200	416,600	269,400	46,800	133,800			588.6	
Output													
③ **Preoxidized Coal**	1,007,800	-	-	44,100	704,300	12,300	68,300	45,000	133,800	800	266 Btu/lb.	268.1	12,641.1
③ **Preoxidizer Vapor**													
Tar + Light Oil (v)	28,300	189	0.8	2,100	23,300	300	2,100	500	x	800	533 Btu/lb.	15.1	452.4
CH_4	1,600	98	0.4	400	1,200	x	x	x	x	"	8,192 Btu/mol	0.8	37.5
C_2H_6	800	26	0.1	200	600	x	x	x	x	"	13,569 "	0.4	17.6
H_2	900	463	1.9	900	x	x	x	x	x	"	5,168 "	2.4	57.0
CO	10,200	364	1.5	x	4,400	x	5,800	x	x	"	5,288 "	1.9	44.3
CO_2	92,500	2,102	8.5	x	25,200	x	67,300	x	x	"	7,784 "	16.4	x
O_2	26,900	842	3.4	x	x	x	26,900	x	x	"	5,494 "	4.6	x
N_2	404,000	14,418	58.3	x	x	404,000	x	x	x	"	5,251 "	75.7	x
H_2S	1,400	41	0.1	100	x	x	x	1,300	x	"	6,525 "	0.3	9.9
H_2O (v)	111,500	6,188	25.0	12,500	x	x	99,000	x	x		1404.7 Btu/lb	156.6	x
Sub-total	678,100	24,731	100.0	16,200	54,700	404,300	201,100	1,800	x			274.2	618.7
Heat Loss												46.3	
Totals	1,685,900	-	-	60,300	759,000	416,600	269,400	46,800	133,800			588.6	

TABLE 13

Mass & Heat Balance
Gasifier
Case III
- See Figure 5

Basis: 1 hour Datum: 60°F $H_2O_{(l)}$

	Pounds	Mols	Mol %	H	C	N	O	S	Ash	MgO·Ca	Temp. °F	ΔH or Cp	ΔH MM Btu	Heat of Combustion HHV MM Btu
Input														
③ Preoxidized Coal	1,007,800	-	-	44,100	704,300	12,300	68,300	45,000	133,800	x	800	266 Btu/lb	268.1	12,641.1
③ Preoxidizer Vapor	678,100	24,731	-	16,200	54,700	404,300	201,100	1,800	x	x	800	(Table 14)	274.2	618.7
④ Air														
O_2	342,000	10,689	20.8	x	x	x	342,000	x	x	x	500	3183 Btu/mol	34.0	x
N_2	1,126,500	40,210	78.3	x	x	1,126,500	x	x	x	x	"	3084 "	124.0	x
H_2O (v)	8,100	418	0.9	900	x	x	7,200	x	x	x	"	1260.2 Btu/lb.	10.2	x
Sub-total	1,476,600	51,317	100.0	900	x	1,126,500	349,200	x	x	x			168.2	x
⑤ Steam	614,700	34,119	-	68,800	x	x	545,900	x	x	x	1200	1609.2 Btu/lb.	989.2	x
⑥ Acceptor														
MgO·CaO	20,268,400	210,253	-	x	x	x	3,364,000	x	x	16,904,400	1900	23.3 Btu/mol-°F	9,014.0	x
Inert	3,576,800	-	-	x	x	x	x	x	3,576,800	x	"	0.255 Btu/lb-°F	1,678.2	x
Sub-total	23,845,200	-	-	x	x	x	3,364,000	x	3,576,800	16,904,400			10,692.2	x
Heat of Reaction														
MgO·CaO to MgO·CaS (1,404 mols x 45,500 Btu/mol)													63.9	
Heat of Combustion (Reactants - Products ex Acceptor)													72.8	
Totals	27,622,400	-	-	130,000	759,000	1,543,100	4,528,500	46,800	3,710,600	16,904,400			12,528.6	
Output														
⑦ Char	298,800	-	-	1,200	166,000	x	x	x	131,600	x	1700	542 Btu/lb	161.9	2,447.3
⑧ Char Loss	16,300	-	-	700	11,400	200	1,100	700	2,200	x	1700	542 Btu/lb	8.8	204.4
⑩ Acceptor														
MgO·CaS	157,900	1,404	-	x	x	x	x	45,000	x	112,900	1700	23.0 Btu/mol-°F	53.0	x
MgO·CaO	20,133,100	208,849	-	x	x	x	3,341,600	x	x	16,791,500	"	23.0 "	7,877.8	x
Inert	3,576,800	-	-	x	x	x	x	x	3,576,800	x	"	0.251 Btu/lb-°F	1,472.3	x
Sub-total	23,867,800	-	-	x	x	x	3,341,600	45,000	3,576,800	16,904,400			9,403.1	x
⑨ Net Gas														
CH_4	3,900	242	0.2	1,000	2,900	x	x	x	x	x	1700	23,134 Btu/mol	5.6	92.6
H_2	92,500	45,907	27.6	92,500	x	x	x	x	x	x	"	11,641 "	534.4	5,654.7
CO	1,081,200	38,602	23.3	x	463,600	x	617,600	x	x	x	"	12,347 "	476.6	4,698.0
CO_2	421,800	9,583	5.8	x	115,100	x	306,700	x	x	x	"	19,290 "	184.8	x
N_2	1,535,900	54,822	33.0	x	x	1,535,900	x	x	x	x	"	12,213 "	669.5	x
NH_3	8,500	501	0.3	1,500	x	7,000	x	x	x	x	"	19,203 "	9.6	82.5
H_2S	1,100	31	x	100	x	x	x	1,000	x	x	"	16,094 "	0.5	7.5
H_2O (v)	294,500	16,345	9.8	33,000	x	x	261,500	x	x	x	"	1885.2 Btu/lb	555.1	x
Sub-total	3,439,400	166,033	100.0	128,100	581,600	1,542,900	1,185,800	1,000	x	x			2,436.1	10,535.3
Heat of Reaction														
Heat of Combustion (1,404 mols H_2S to CaS)													339.7	
Recycle Gas Cooling (1560 mols 1700°F to 1600°F)													1.4	
Heat Loss													177.6	
Totals	27,622,300	-	-	130,000	759,000	1,543,100	4,528,500	46,700	3,710,600	16,904,400			12,528.6	

TABLE 16

Mass & Heat Balance
Regenerator
Case III

- See Figure 5

Basis: 1 hour Datum: 60°F H_2O(l)

	Pounds	Mols	Mol %	Elemental Balance, Pounds							Temp. °F	ΔH or Cp	ΔH MM Btu	Heat of Combustion HHV MM Btu
				H	C	N	O	S	Ash	MgO·Ca				
Input														
⑦ Char	298,800	-	-	1,200	166,000	x	x	x	131,600	x	1700	542 Btu/lb	161.9	2,447.3
⑪ Lift Gas														
CO_2	54,900	1,248	19.4	x	15,000	x	39,900	x	x	x	280	2072 Btu/mol	2.6	x
N_2	136,400	4,870	75.9	x	x	136,400	x	x	x	x	"	1534 "	7.5	x
H_2S	200	6	0.1	x	x	x	x	200	x	x	"	-	x	1.3
H_2O (v)	5,300	296	4.6	600	x	x	4,700	x	x	x	"	1158.4 Btu/lb	6.2	x
Sub-total	196,800	6,420	100.0	600	15,000	136,400	44,600	200	x	x			16.3	1.3
⑫ Air														
O_2	474,700	14,833	20.8	x	x	x	474,700	x	x	x	315	1816 Btu/mol	26.9	x
N_2	1,563,300	55,800	78.3	x	x	1,563,300	x	x	x	x	"	1779 "	99.3	x
H_2O (v)	11,200	621	0.9	1,300	x	x	9,900	x	x	x	"	1174.5 Btu/lb	13.1	x
Sub-total	2,049,200	71,254	100.0	1,300	x	1,563,300	484,600	x	x	x			139.3	x
⑬ Acceptor														
MgO·CaS	157,500	1,401	-	x	x	x	x	44,900	x	112,600	1700	23.0 Btu/mol-°F	52.8	x
MgO·CaO	20,092,800	208,431	-	x	x	x	3,334,900	x	x	16,757,900	"	23.0 "	7,862.0	x
Inert	3,569,600	-	-	x	x	x	x	x	3,569,600	x	"	0.251 Btu/lb-°F	1,469.4	x
Sub-total	23,819,900	-	-	x	x	x	3,334,900	44,900	3,569,600	16,870,500			9,384.2	x
⑭ Make-up Acceptor														
$MgCO_3$ $CaCO_3$	77,600	421	-	x	10,100	x	33,700	x	x	33,800	60	-	x	x
Inert	7,200	-	-	x	x	x	x	x	7,200	x	"	-	x	x
Sub-total	81,800	-	-	x	10,100	x	33,700	x	7,200	33,800			x	x
Heat of Reaction														
MgO CaS to MgO CaO (1,401 mols x 196,600 Btu/mol)													275.4	
Heat of Combustion (Reactants - Products ex Acceptor)													2,212.4	
Totals	26,449,500	-	-	3,100	191,100	1,699,700	3,897,800	45,100	3,708,400	16,904,300			12,189.5	
Output														
⑮ Char	148,200	-	-	x	16,600	x	x	x	131,600	x	1900	496 Btu/lb.	73.6	236.2
⑥ Acceptor														
MgO CaO	20,268,400	210,253	-	x	x	x	3,364,000	x	x	16,904,400	1900	23.3 Btu/mol-°F	9,014.0	x
Inert	3,576,800	-	-	x	x	x	x	x	3,576,800	x	"	0.255 Btu/lb-°F	1,678.2	x
Sub-total	23,845,200	-	-	x	x	x	3,364,000	x	3,576,800	16,904,400			10,692.2	x
⑯ Gas														
CO_2	639,300	14,525	18.6	x	174,500	x	464,800	x	x	x	1900	22,012 Btu/mol	319.7	x
N_2	1,699,700	60,670	77.7	x	x	1,699,700	x	x	x	x	"	13,837 "	839.5	x
SO_2	90,100	1,407	1.8	x	x	x	45,000	45,100	x	x	"	22,247 "	31.3	x
H_2O (v)	27,200	1,510	1.9	3,100	x	x	24,100	x	x	x	"	1997.1 Btu/lb.	54.3	x
Sub-total	2,456,300	78,112	100.0	3,100	174,500	1,699,700	533,900	45,100	x	x			1,244.8	x
Heat of Reaction														
$MgCO_3$ $CaCO_3$ to MgO CaO (421 mols x 127,400 Btu/mol)													53.8	
Heat Loss													125.3	
Totals	26,449,500	-	-	3,100	191,100	1,699,700	3,897,800	45,100	3,708,400	16,904,400			12,189.5	

TABLE 17

Mass & Heat Balance
Reductor
Case III

- See Figures 5 & 8

Basis: 1 hour Datum: 60°F $H_2O_{(1)}$

	Pounds	Mols	Temp. °F	ΔH or Cp	ΔH MM Btu
Input					
⑯ **Regenerator Gas**					
CO_2	639,300	14,525	695	6556 Btu/mol	95.2
N_2	1,699,700	60,670	695	4485 Btu/mol	272.1
SO_2	90,100	1,407	695	6859 Btu/mol	9.6
H_2O (v)	27,200	1,510	695	1353.2 Btu/lb	36.8
Sub-total	2,456,300	78,112			413.7
⑰ **Gasifier Gas**					
CH_4	100	8	128	--	--
H_2	3,100	1,526	128	470 Btu/mol	0.7
CO	35,900	1,283	128	473 Btu/mol	0.6
CO_2	14,000	318	128	616 Btu/mol	0.2
N_2	51,100	1,822	128	473 Btu/mol	0.9
H_2O (v)	9,800	543	128	1088.9 Btu/lb	10.7
Sub-total	114,000	5,500			13.1
⑱ **Stripping Gas**					
CO_2	2,900	65	120	--	--
N_2	7,400	265	120	417 Btu/mol	0.1
H_2S	100	3	120	--	--
H_2O (v)	400	24	120	1085.6 Btu/lb	0.5
Sub-total	10,800	357			0.6
Heat of Reaction: Heat of Combustion (Reactants - Products)					120.1
Totals	2,581,100	83,969			547.5
Output					
⑭ **Reductor Gas**					
CH_4	100	8	800	8192 Btu/mol	0.1
CO_2	711,800	16,173	800	7784 Btu/mol	125.9
N_2	1,758,200	62,757	800	5251 Btu/mol	329.5
H_2S	21,200	623	800	6525 Btu/mol	4.0
COS	1,100	18	800	8510 Btu/mol	0.2
SO_2	20,600	322	800	8123 Btu/mol	2.6
S_2 (g)	14,300	224	800	6144 Btu/mol	1.4
H_2O (v)	53,700	2,982	800	1404.7 Btu/lb	75.4
Sub-total	2,581,000	83,107			539.1
Heat Loss					8.4
Totals	2,581,000	83,107			547.5

TABLE 18

Mass & Heat Balance
Claus Reactor
Case III
- See Figures 5 & 8

Basis: 1 hour Datum: 60°F $H_2O_{(1)}$

	Pounds	Mols	Temp. °F	ΔH or Cp		ΔH MM Btu
Input						
⑲ Reductor Gas						
CH_4	100	8	402	3,265	Btu/mol	x
CO_2	711,800	16,173	"	3,317	"	53.6
N_2	1,758,200	62,757	"	2,390	"	150.0
H_2S	21,200	623	"	2,868	"	1.8
COS	1,100	18	"	3,682	"	0.1
SO_2	20,600	322	"	3,502	"	1.1
S_8	14,300	56	"	13,400	"	0.8
H_2O (v)	53,700	2,982	"	1214.6	Btu/lb.	65.2
Sub-total	2,581,000	82,939				272.6
Heat of Reaction						
$268\ SO_2 + 18\ COS + 518\ H_2S \longrightarrow$						
$100.5\ S_8 + 18\ CO_2 + 518\ H_2O_{(1)}$						22.6
Totals	2,581,000	82,939				295.2
Output						
⑳ Claus Gas						
CH_4	100	8	410	3,355	Btu/mol	x
CO_2	712,600	16,191	"	3,403	"	55.1
N_2	1,758,200	62,757	"	2,447	"	153.6
H_2S	3,600	105	"	2,939	"	0.3
SO_2	3,400	53	"	3,592	"	0.2
S_8	40,100	156.5	"	13,729	"	2.1
H_2O (v)	63,000	3,500	"	1218.3	Btu/lb	76.8
Sub-total	2,581,000	82,770				288.1
Heat Loss						7.1
Totals	2,581,000	82,770				295.2

4. Economic Evaluation - Gasification Section

Basis for Cost Estimate

All of the cost estimates in this study are based on earlier Consol studies on the CO_2 Acceptor Process carried out for the Office of Coal Research(1d) updated to present-day cost indices. While the cost of individual pieces of equipment may be estimated by this method with reasonable accuracy, the estimate of indirect construction costs (e.g., piping, instrumentation, etc.) is of necessity quite imprecise. Therefore, the absolute accuracy of these cost estimates is uncertain. Differences in costs between the various cases are probably reasonably accurate since the same method was used in all instances.

To determine the total investment costs, the following assumptions were made, and are common to all the gasification cases as well as to the combined cycle power system described in Section III-B-5, page 50.

1. Minemouth site.
2. January 1971 costs.*
3. Begin construction July 1971.
4. Operation January 1976.
5. Escalation at 7-1/2% average of labor and material.
6. Interest during construction at 7-1/2% simple interest on cash flow.
7. Cooling towers not included.

* Costs to be quoted later for the United Aircraft Research Laboratories power cycle are as of January, 1970.

Escalation of direct operating costs in all cases was taken at the basic rate of 7-1/2% per year average for labor and material, subject to the construction schedule outlined above.

The coal costs used in this study represent Consol's estimate for the minemouth cost of coal from a new mine in West Virginia to be opened prior to operation of the power plant.

Basic Pressure Case I

Based on the process design for this case described previously, and the detailed material and heat balances reported in Tables 4 through 9, a cost estimate for each section of the plant was evolved as shown in Appendix A. Tables A-1 to A-3 in Appendix A show the ISBL (inside battery limits) cost estimates for coal preparation, gasification, and sulfur recovery-solids disposal, respectively. To these costs must be added off-site facilities, utility costs, electrical substations, cooling water towers, distribution piping, and boiler feed water treating.

The installed plant cost is $81,000,000, not including interest and escalation, and a breakdown by section is shown in Table 19. This is the installed cost for the basic process scheme wherein the product gas is available under pressure at 1300°F without water scrubbing. The investment cost

TABLE 19

Investment Summary
Case I

Section	Coal Preparation	Gasification	Sulfur Recovery-Solids Disposal	Totals
Utility Requirements				
Electricity, KW	4,200	12,400	1,000	17,600
Cooling Water, GPM	x	17,300	3,700	21,000
Boiler Feed Water, GPM	x	1,463	x	1,463
Investment, $1000				
Erected Cost (ISBL)	5,000	56,000	11,000	72,000
Off-sites & Utilities				
Off-sites	400	4,300	900	5,600
Electrical	500	1,500	100	2,100
Cooling Water	x	600	100	700
Boiler Feed Water	x	600	x	600
Sub-Total (OSBL)	900	7,000	1,100	9,000
Installed Plant Cost	5,900	63,000	12,100	81,000
			Escalation	15,200
				96,200
			Interest During Construction	16,200
			Total	$112,400

includes \$13 MM (installed cost, including accessory electrical equipment) for motor-driven compressors to supply the process air. The compressors are more properly charged to the power cycle in a case of a combined cycle power station in order to give the "base" cost of the gasification system only. This base cost then becomes \$68 MM (\$94 MM including escalation and interest during construction). Costs associated with supply of process air for various modes of integration with a power station are shown in Sections III-B-5 and 6. If water scrubbing to eliminate alkalis and entrained dust is required, and the product gas is reheated to about 1100°F, the installed base plant cost is increased to \$94 MM. Most of the \$13 MM (\$18 MM with escalation and interest during construction) incremental cost is associated with the very large gas-gas heat exchangers which are required.

Direct operating costs for the basic flow scheme were estimated as shown in Table 20.

Alternate Pressure Case II

The alternate pressure case yields almost the same quantity of product gas at the same conditions of 1300°F and 206 psia. The higher investment cost (\$4 MM) shown in Table 21 for Case II is largely the result of increased vessel costs in the sulfur recovery section and the need for stainless steel-lined equipment to cope with the corrosive H_2SO_3 stream. The somewhat higher direct operating costs shown in Table 21 stem from increased operating labor and power costs for compression of the air and gas streams in the sulfur recovery section.

Atmospheric Pressure Case III

To fulfill the requirements of Phase I under the contract, the same economic evaluation of this case was prepared as discussed for Case I.

The detailed ISBL estimates are shown in Tables A-4 to A-6 in Appendix A. The summarized cost for the installed plant is given in Table 22, and the direct operating costs in Table 23.

5. Integration with Combined Cycle Power Systems

Basis for Evaluation

The cost of producing power from low-sulfur boiler fuel will be relatively high when the CO_2 Acceptor Process is used simply to produce clean fuel gas for an existing boiler. Improved economics and station heat rates can be achieved by integration of the gasification section with combined cycle power systems. This section gives the design basis used for evaluation of such an integrated system.

Two studies, sponsored by OAP, were made by Westinghouse Research Laboratories (OAP Contract No. CPA-70-9) and by United Aircraft Research Laboratories(19) in which designs and cost estimates for advanced power cycles were performed. In Germany, STEAG is building a 170 MW combined cycle power station which uses the Lurgi pressure gasification process to provide low-Btu fuel gas.

TABLE 20

Direct Operating Cost Summary, Excluding Coal Cost
Case I

Basis: 70% Plant Operating Factor

Section	Coal Preparation	Gasification	Sulfur Recovery-Solids Disposal	Totals
Direct Operating Labor, Men/Shift	3	10	6	19
Direct Operating Cost	$1000/Year			
1. Operating Labor at $40,000/man/shift/yr	120	400	240	760
2. Maintenance Labor at 1.6% Investment	94	1,008	194	1,296
3. Direct Supervision 15% of 1. + 2.	32	211	65	308
4. Indirect Overhead 50% of 1. + 2. + 3.	123	809	250	1,182
5. Payroll Overhead 15% of 1. + 2. + 3. + 4.	55	364	113	532
6. Maintenance Material at 2.4% Investment	142	1,512	290	1,944
7. Miscellaneous Supplies 15% of Maintenance Material	21	227	44	292
8. Utilities				
Electricity at 7.5 mills/KWH	194	570(1)	46	810
Cooling Water at 2¢/1000 gal	--	128	27	155
BFW at 25¢/1000 gal	--	135	--	135
Chemicals & Catalyst	--	830	71	901
TOTALS	781	6,194	1,340	8,315
			Escalation	1,560(2)
			Total	9,875

---

(1) Ex. main air compressor duty.

(2) Direct operating costs shown here are investment-sensitive (Items 2 through 7). Therefore, these costs are escalated in the same manner used for the plant investment shown in Table 19, i.e., the average escalation over the 5-year period, 1971-1976, at a rate of 7.5%/yr for labor and materials.

TABLE 21

Economic Comparison - Cases I and II

	Case I	Case II
Installed Plant Cost	$81,000,000	$85,000,000
Annual Direct Operating Costs (ex Coal)	$ 8,315,000/yr	$ 8,776,000/yr

---

TABLE 22

Investment Summary
Case III

Section	Coal Preparation	Gasification	Sulfur Recovery-Solids Disposal	Totals
Utility Requirements				
Electricity, KW	4750	6200	1350	12,300
Cooling Water, GPM	90	45,680	1160	46,930
Boiler Feed Water, GPM	--	1,310	--	1,310
Investment, $1000				
Erected Cost (ISBL)	4,500	76,500	10,000	91,000
Offsites & Utilities				
Offsites	400	6,100	800	7,300
Electrical	600	700	200	1,500
Cooling Water	--	1,600	--	1,600
Boiler Feed Water	--	600	--	600
Sub-total (OSBL)	1,000	9,000	1,000	11,000
Installed Plant Cost	5,500	85,500	11,000	102,000
			Escalation	19,200
				121,200
			Interest During Construction	20,500
			Total	141,700

TABLE 23

Direct Operating Cost Summary,
Excluding Coal Cost
Case III

Basis: 70% Plant Operating Factor

Section	Coal Preparation	Gasification	Sulfur Recovery-Solids Disposal	Totals
Direct Operating Labor, Men/Shift	3	10	6	19
Direct Operating Costs	$1000/Year			
1. Operating Labor at $40,000/man/shift/yr	120	400	240	760
2. Maintenance Labor at 1.6% Investment	88	1,368	176	1,632
3. Direct Supervision 15% of 1. + 2.	31	265	63	359
4. Indirect Overhead 50% of 1. + 2. + 3.	120	1,016	240	1,376
5. Payroll Overhead 15% of 1. + 2. + 3. + 4.	54	457	108	619
6. Maintenanct Material at 2.4% Investment	132	2,052	264	2,448
7. Miscellaneous Supplies 15% of Maintenance Material	20	308	39	367
8. Utilities				
Electricity at 7.5 mills/KWH	219	285	63	567
Cooling Water at 2¢/1000 gal	1	336	9	346
BFW at 25¢/1000 gal	--	121	--	121
Chemicals & Catalyst	--	910	105	1,015
TOTALS	785	7,518	1,307	9,610
			Escalation	1,800
			Total	11,410

In this section, integration of the Case I fuel gas with two advanced power cycles is considered. It is emphasized that no detailed designs and cost estimates for power station components were made. Results of the Westinghouse and UARL studies, information obtained from Lurgi,(8) and costs for conventional coal-fired stations presented by United Engineers and Constructors(20) provided the bases for estimation.

For each power cycle, two versions of the Case I fuel gas product were used. In one, the assumption was made that particulate matter and alkali in the product gas, after cooling to 1300°F, could be reduced sufficiently by high pressure drop cyclones to allow sustained operation of gas turbine engines and expanders. In the other, the assumption was made that wet scrubbing would be required for adequate cleanup. In this instance the gas is wet scrubbed at the saturation temperature (~240°F) corresponding to its steam content. The gas then is reheated in a heat exchanger to 1100°F. Also, ammonia formed from the coal nitrogen in the gasifier was assumed to be removed from the fuel gas during scrubbing.

Supercharged Boiler Cycle

In this power cycle fuel gas from the CO_2 acceptor gasifier is burned at elevated pressures in a boiler which provides steam for power generation by a conventional steam cycle. Hot gas from the boiler is then passed through an expander which drives the compressor for combustion air and a generator which produces additional electrical power. The expander exhaust gas is used to preheat the boiler feed water before final discharge to the stack. This cycle is used in the Westinghouse study (high-pressure fluidized bed boiler) and in STEAG's Kellermann station (high-pressure Lurgi fuel gas).

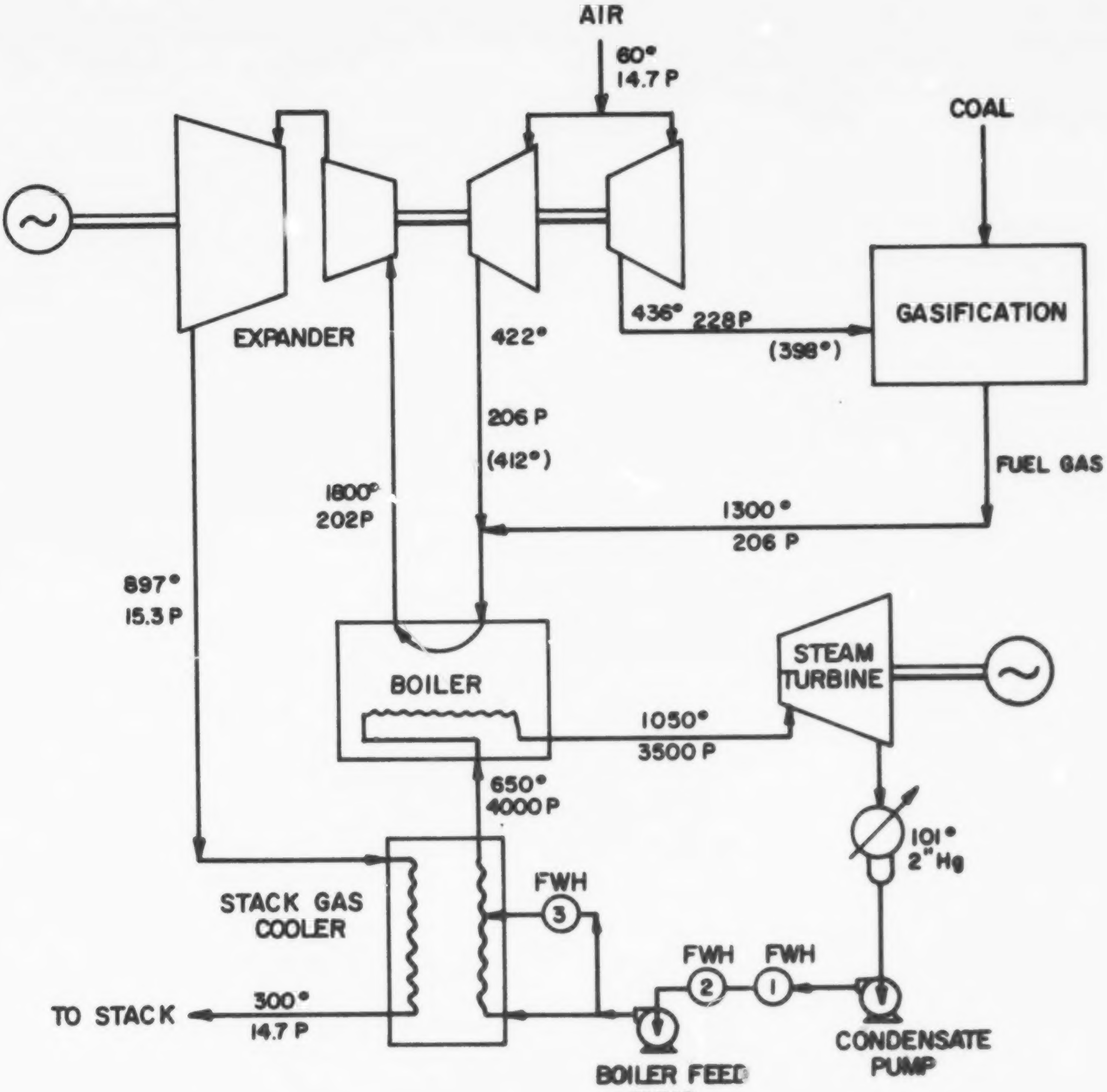
Figure 9 is a schematic diagram showing integration of Consol's Case I gasification process with the supercharged boiler cycle. Flows and power outputs are summarized in Table 24. For the steam cycle, steam conditions of 3500 psia/1050°F/1050°F were chosen on the basis of high efficiency. Detailed design and cost estimation would be required to show whether this cycle is justified, compared with a lower cost subcritical cycle. Gross output of the expander was taken as 91% of the isentropic output for the particular working fluid used. Air compression duties were calculated at 89% of the polytropic efficiency. Thermodynamic properties of dry air were taken from the NBS tables.(9)

On the basis of available technology, using cooled bleed air for blade cooling, the expander inlet gas temperature was taken at 1800°F. No specification is made as to the unit sizes of the expanders and steam turbines. By the nature of the gasification process (12 gasifiers and 4 regenerators, as shown in Figure 3), multiple units are envisioned.

The assumption was made that complete combustion of the fuel gas can be achieved in the supercharged boiler with 5% excess air. At this level of excess air and with the use of two-stage combustion(21) of the low-Btu fuel gas, formation of nitrogen oxides should be drastically reduced from the levels associated with conventional PF boilers.

FIGURE 9

SCHEMATIC FLOW DIAGRAM
CASE I GASIFICATION SYSTEM
SUPERCHARGED BOILER



AIR
60°
14.7 P
COAL
EXPANDER
422°
436° 228P
(398°)
GASIFICATION
206 P
(412°)
FUEL GAS
1800°
202P
1300°
206 P
897°
15.3 P
BOILER
STEAM TURBINE
1050°
3500 P
650°
4000 P
101°
2" Hg
STACK GAS COOLER
FWH
3
FWH
2
FWH
1
TO STACK
300°
14.7 P
BOILER FEED PUMP
CONDENSATE PUMP

Heat Source For FeedWater Heaters:

① From Air Compressor Intercooler

② Multistage, By Steam Extracted From LP Turbine.

③ Multistage, " " " " IP "

° Fahrenheit
P = PSIA

TABLE 24

Power Cycles

Case I Gasifier
Supercharged Boiler Power Cycle

Method of Fuel Gas Cleanup (particulate and alkali)	Hot	Wet Scrubbing
Gas Cycle		
Expanders		
Inlet Gas		
Mols/hr	414,900	412,400
Temperature, °F	1800	1800
Pressure, psia	202	196
Exhaust Gas		
Temperature, °F	897	905
Pressure, psia	15.3	15.3
Air Compressors		
Combustion Air		
Mols/hr	216,900	215,100
Exit temperature, °F(1)	422	418
Exit pressure, psia	206	200
Process Air		
Mols/hr	136,500	136,500
Exit temperature, °F	436	436
Exit pressure, psia	228	228
Bleed air for cooling (150°F), % of total airflow	4.0	4.0
Busbar Power Output, MW(2)	513	494
Steam Cycle		
Steam Turbines		
Steam Flow, MM lb/hr		
Main steam	6.017	5.578
Reheat steam	5.950	5.520
Extracted for FW heating	1.464	1.222
Throttle temperature, °F	1050	1050
Throttle pressure, psia	3500	3500
Reheat temperature, °F	1050	1050
Reheat pressure, psia	620	620
Condenser temperature, °F	101	101
Condenser pressure, inches Hg	2.0	2.0
Boilers		
Fuel Gas		
Mols/hr	234,980	234,470
Temperature, °F	1300	1100
Pressure, psia	206	200
Combustion Air		
Mols/hr	216,900	215,100
Temperature, °F	412	408
Pressure, psia	206	200
Products of Combustion (5% excess air)		
Mols/hr	414,900	412,400
Temperature, °F	1800	1800
Pressure, psia	202	196
Feedwater		
MM lb/hr	6.017	5.578
Temperature, °F	650	650
Pressure, psia	4000	4000
Stack Gas		
Mols/hr	429,600	427,000
Temperature, °F	300	300
Pressure, psia	14.7	14.7
Busbar Power Output, MW(3)	1049	983

(1) One stage of intercooling to 120°F.
(2) 98% of expander power output minus total compressor duty.
(3) After allowance of 3.9% for auxiliaries, heat and generator losses.

Exhaust Gas Cycle with 2800°F Turbine Inlet Temperature

This cycle is used in the UARL study.(3,19) Fuel gas is burned with sufficient excess air in the combustor of a gas turbine engine to provide inlet gas to the compressor turbine at 2800°F. The power turbine drives an electric generator. The hot exhaust gas from the power turbine provides steam in a heat recovery boiler for power generation by a conventional steam cycle. No fuel is burned in the boiler and the feed water is not preheated.

Figure 10 is a schematic diagram showing integration of Consol's Case I with the exhaust gas cycle. Flows and power outputs are shown in Table 25.

Since the fuel gas composition and sensible heat content are appreciably different from those used in the UARL study, some modifications were required to adapt the UARL estimate for use in this study. The gross power outputs of the compressor and power turbines were calculated, using 91% of isentropic efficiency for the particular working fluid used in Case I. Air compression duties, steam conditions, and steam cycle efficiencies are on the same basis as in the UARL study.

In Case I, the fuel gas is produced at somewhat higher pressure than the 190 psia level used by UARL. The schematic diagram shows an expander, which produces additional power, and which brings the fuel gas to the same pressure as used by UARL. In practice, the gas turbine engine would be modified to accommodate the higher inlet pressure. However, to keep the same pressure ratios in the gas turbine engine as used by UARL, the expander is necessary.

Economics

The capital costs for the gasification system are taken from Table 19. The costs for the power stations, unit costs ($/KW of installed capacity) were obtained from four sources, as outlined below:

	Power Cycle	
Category	Supercharged Boiler	Exhaust Gas
Structures and Improvements	UE	U
Boiler Plant	L	U
Steam Turbine - Generator	UE	U
Gas Turbine - Generator	W2	U
Stack Gas Cooler	W1	-
Switchyard and Station Equipment	UE	UE
Accessory Electrical Equipment and Miscellaneous	UE	U
Indirect Construction Costs	UE*	U
Land and Contingency	UE**	

Key:

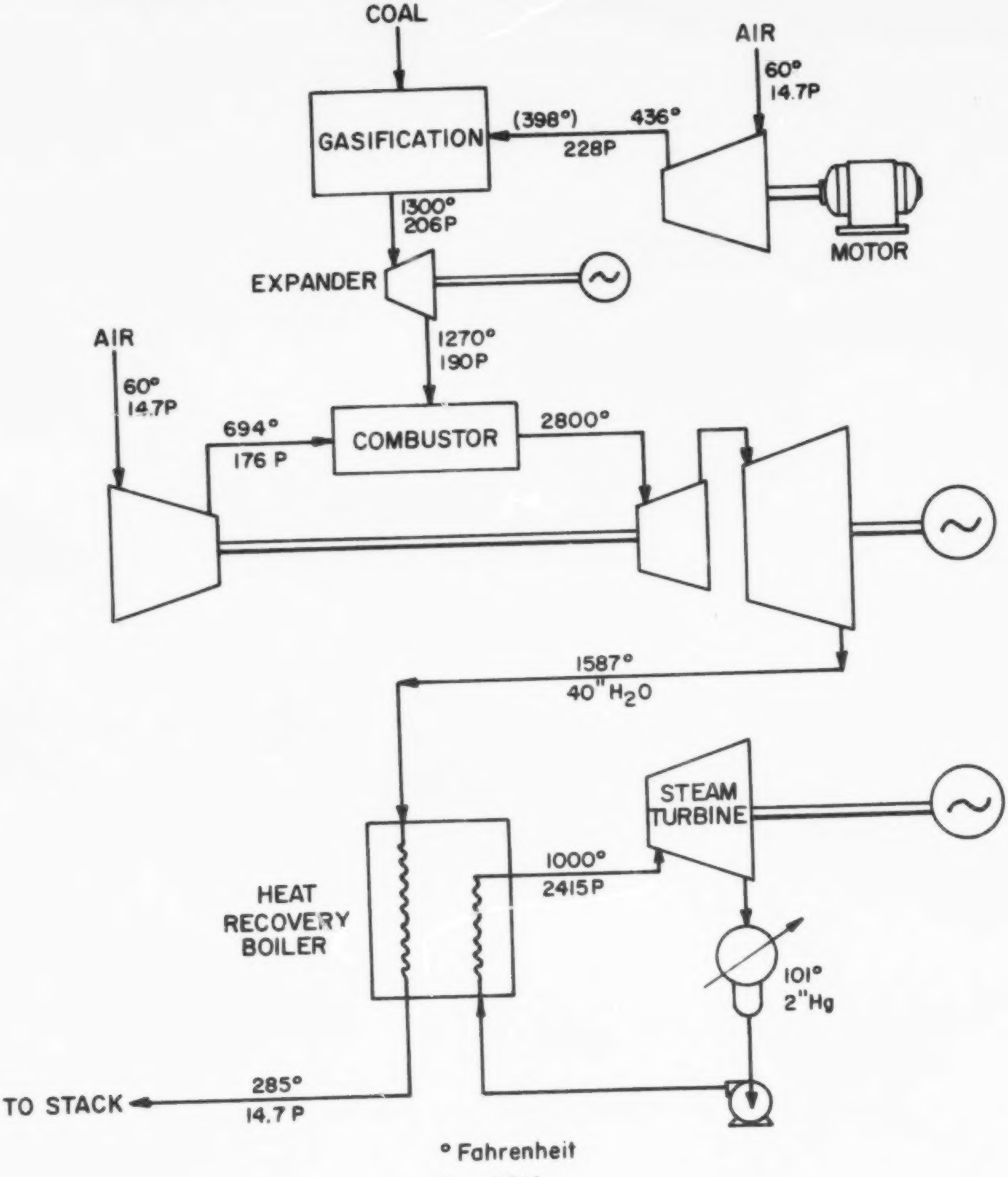
W1 Westinghouse, Reference 10.
W2 Westinghouse, Reference 10a.
UE United Engineers and Constructors, Reference 20.
L Lurgi, Reference 8.
U UARL, Reference 3, 19.

* Indirect construction costs taken as 7.4% of direct cost, as shown in Reference 20.

** Contingency taken as 6.4% of direct + indirect costs, as shown in Reference 20.

FIGURE 10

SCHEMATIC FLOW DIAGRAM
CASE I GASIFICATION SYSTEM - EXHAUST GAS CYCLE
2800°F TURBINE INLET TEMPERATURE


COAL
AIR
60°
14.7P
GASIFICATION
(398°)
436°
228P
MOTOR
1300°
206P
EXPANDER
1270°
190P
AIR
60°
14.7P
694°
176 P
COMBUSTOR
2800°
1587°
40" H_2O
STEAM
TURBINE
1000°
2415P
HEAT
RECOVERY
BOILER
101°
2" Hg
TO STACK
285°
14.7 P

° Fahrenheit

P = PSIA

TABLE 25

Power Cycles

Case I - Gasification System
Exhaust Gas Power Cycle with 2800°F Turbine Inlet Temperature

Method of Fuel Gas Cleanup (particulate and alkali)	Hot	Wet Scrubbing
Gas Cycle		
Engines		
Compressor Pressure Ratio	12	12
Inlet Gas		
Mols/hr	565,100	540,000
Temperature, °F	2800	2800
Exhaust Gas		
Mols/hr	599,400	572,000
Temperature, °F	1587	1590
Bleed air for cooling (200°F), % of total airflow	8.54	8.54
Compressor Airflow, mols/hr		
Combustion air	367,100	343,000
Bleed air	34,300	32,100
Compressor outlet temperature, °F	694	694
Fuel Gas Expanders		
Inlet pressure, psia	206	200
Inlet temperature, °F	1300	1100
Outlet pressure, psia	190	190
Outlet temperature, °F	1270	1081
Busbar Power Output, MW		
Engines (1)	1310	1266
Expanders	17	10
Total	1327	1276
Steam Cycle		
Steam Turbines (2)		
Throttle temperature, °F	1000	1000
Throttle pressure, psig	2400	2400
Throttle flow, MM lb/hr	4272	3944
Reheat temperature, °F	1000	1000
Reheat pressure, psig	569	569
Reheat flow, MM lb/hr	4218	3894
Heat Recovery Boilers		
Inlet Gas		
Heat content, MM Btu/hr	7210	6690
Temperature, °F	1587	1590
Pressure, psia	16.17	16.17
Stack Gas		
Heat content, MM Btu/hr	971	930
Temperature, °F	285	285
Pressure, psia	14.7	14.7
Net Boiler Duty, MM Btu/hr	6239	5760
Busbar Power Output, MW (3)	726	670

(1) 98% of gross power output minus compressor duty.
(2) Same steam conditions as used by United Aircraft Research Laboratories, Reference 3.
(3) At same ratio of power output/net boiler duty as used by United Aircraft Research Laboratories.

Supercharged boiler costs obtained from Lurgi for the Kellermann plant were arbitrarily increased by 50% to compensate for addition of reheater surface and for the higher steam pressure. The Kellerman plant uses a modest steam cycle: 1900 psi/1000°F/no reheat.

Table 26 shows the summarized power station investment costs and includes a comparison with a modern conventional PF station equipped with Consol's Formate Process(17) for stack gas scrubbing. Investment and operating costs for stack gas scrubbing are from unpublished work by Consol. The Formate Process for stack gas scrubbing was chosen to put the sulfur recovery systems on a comparable basis, i.e., sulfur is recovered in all cases as elemental sulfur. The major item in the operating cost of the Formate Process is the cost of fuel gas required to provide reducing gas for the regeneration section. Originally(17) natural gas was used as fuel at 40¢/MM Btu. Since that time a severe shortage of natural gas has developed, and it was necessary for this study to increase the fuel costs for the regeneration section from 40¢ to 80¢/MM Btu.

Total costs for a conventional PF station were presented by United Engineers and Constructors in reference 20. Breakdown into the individual Federal Power Commission account numbers was on the basis of a previous study by UEC which was made for Westinghouse in 1965.(11)

In the supercharged boiler systems, process air for the gasification plant is supplied by compressors driven by the power station expanders. Thus, the investment associated with process air are included in the entries for "Gas Turbine - Generator" in Table 26. In the exhaust gas systems, the process air is supplied by motor-driven compressors, as in the UARL study. Thus, the gasification plant investment corresponds exactly to that shown in Table 19.

Table 27 shows the summarized power costs for the same stations which are in Table 26. Direct operating costs for the power plants shown in Table 27 attempt to reflect the higher maintenance costs which probably will be associated with advanced cycle gas turbines and expanders and to reflect the elimination of coal and ash handling costs in the combined cycle stations. The direct operating costs for the conventional coal-fired station were based on annual Federal Power Commission data for modern, efficient stations.

As Table 27 shows, there is considerable incentive to develop a hot gas clean-up process which operates at about 1300°F. The increased costs due to wet gas scrubbing can be reduced somewhat over those shown in the table if the heat rejected to cooling water in the scrubber circuit (~450 MM Btu/hr in Case I) could be utilized in the power station. This possibility was not investigated.

6. Low-Sulfur Fuel Gas for Conventional Stations

Introduction

Integration of the low-sulfur gas process with advanced cycle power stations appears to be attractive for long-term requirements for power generation.

TABLE 26

Summarized Power System Investment Costs
1976 Operation. See page 48 for Plant Cost Basis

System	Conventional Coal Fired With Stack Gas Scrubbing	Case I - Gasification With: Supercharged Boiler, Hot	Case I - Gasification With: Supercharged Boiler, Wet Scrubbing	Case I - Gasification With: Exhaust Gas Cycle With 2800° F Turbine Inlet Temperature, Hot	Case I - Gasification With: Exhaust Gas Cycle With 2800° F Turbine Inlet Temperature, Wet Scrubbing
Fuel Gas Cleanup (particulate and alkali)	--				
Busbar Power, MW	1000[1]	1562	1477	2053	1946
Air Compression, MW	--	--	--	168	168
Miscellaneous Drives, MW	--	18	18	18	18
Net Power, MW	1000	1544	1459	1867	1760
Sulfur Removal, % (as elemental sulfur)	90	96			
Overall Station Heat Rate, Btu/Net KWH	8900	8930	9450	7390	7840
Busbar Power, % from					
Steam Cycle	100	67.2	66.5	35.3	34.4
Gas Cycle	0	32.8	33.5	64.7	65.6
System Investment					
Power Station, $/KW Busbar Power					
Year for Cost Basis	1971	1971	1971	1970	1970
Structures and Improvements	13.2	9.5	9.5	7.7	7.7
Boiler Plant	75.3	16.8	16.6	13.2	12.8
Steam Turbine - Generator	30.3	20.3	20.1	13.5	13.2
Gas Turbine - Generator	--	18.5	19.3	19.8	19.9
Stack Gas Cooler	--	9.0	9.0	--	--
Switchyard and Station Equipment	9.7	9.7	9.7	9.7	9.7
Acc. Elect. Equip. and Misc.	8.8	8.8	8.8	7.2	7.2
Land	.4	.4	.4	.2	.2
Indirect Construction Costs	10.2	6.9	6.9	10.6	10.6
Contingency	9.5	6.4	6.4		
Sulfur Removal	22.4[2]	[3]	[3]	[3]	[3]
Total Capitalized Cost	179.8	106.3	106.7	81.9	81.3
Escalation	33.7	19.9	20.0	23.0	22.9
	213.5	126.2	126.7	104.9	104.2
Interest During Construction	36.0	21.3	21.4	17.7	17.6
Total	249.5	147.5	148.1	122.6	121.8
$/KW Net Power	249.5	149.2	150.0	134.8	134.7
Gasification Plant, MM$ Inv.					
Base Plant	--	94.4	94.4	94.4	94.4
Air Compression	--	--	--	18.0	18.0
Wet Scrubbing	--	--	18.0	--	18.0
Total		94.4	112.4	112.4	130.5
$/KW Net Power	--	61.1	77.0	60.2	74.1
Total System Cost, $/KW Net Power	249.5	210.3	227.0	195.0	208.8

---

[1] After subtraction of power for auxiliaries.
[2] Via Consol's Formate Process. Reference 17 - Investment does not include separate scrubber for removal of residual particulate matter prior to SO_2 removal, if required.
[3] In gasification plant.

TABLE 27

Power Costs
1976 Operation. See page [illegible] for Plant Cost Basis

Basis:

Coal
at 40¢/MM Btu
4.3% Sulfur Content
12,700 Btu/lb

15% Capital Charges
6132 hrs/yr = 70% plant factor
$15/LT Sulfur Credit

System	Conventional Coal Fired With Stack Gas Scrubbing	Case 1 - Gasification With			
		Supercharged Boiler		Exhaust Gas Cycle With 2800°F Turbine Inlet Temperature	
Fuel Gas Cleanup (Particulate and alkali)	--	Hot	Wet Scrubbing	Hot	Wet Scrubbing
Heat Rate, Btu/KWH					
Busbar Power	8900(1,2)	8838	9347	6727	7099
Net Power	8900(2)	8930	9450	7390	7840
Investment, $/KW Net Power					
Gasification	31.1(3)	61.1	77.0	60.2	74.1
Power Plant	218.4	149.2	150.0	134.8	134.7
Total Station	249.5	210.3	227.0	195.0	208.8
	Mills/KWH Busbar Power				
Coal	3.56	3.54	3.74	2.69	2.84
Gasification					
Direct Operating Cost	.77(3)	.94	1.09	.78	.90
Capital Charges	.76(3)	1.48	1.86	1.34	1.64
Power Cost for Air Compression	--	--	--	.74	.78
Sub Total Gasification	1.53(3)	2.42	2.95	2.86	3.32
Power Plant					
Direct Operating Cost	.48	.48	.48	.59	.59
Capital Charges	5.34	3.61	3.62	3.00	2.98
Sulfur Credit	-.15	-.17	-.18	-.13	-.14
Total	10.76	9.88	10.61	9.01	9.59

(1) After subtraction of power for auxiliaries.
(2) Based on coal burned in boiler.
(3) Includes investment for sulfur removal via Consol's Formate Process. Reference 17 - Investment for separate scrubber for removal of residual particulate, which may be required, is not included. 640 Btu/KWH fuel equivalent is required to supply reductant for sulfur recovery, fuel for fired heaters and steam and power to drive fans and pumps in SO_2 and sulfur recovery sections. Major purchased fuel is natural gas which is included in direct operating cost at 80¢/MM Btu.

Existing stations, now faced with restrictions on sulfur emissions and with future restrictions on nitrogen oxide emissions require a shorter-term approach. Successful development of the CO_2 Acceptor Process through Phases II and III of the OAP contract would provide a practical alternative for sulfur cleanup in the event that a stack gas treatment process would not be feasible or practical.

Use of fuel gas produced by the CO_2 Acceptor Process in a conventional boiler also potentially offers the following advantages over stack gas clean-up processes:

a. Lower nitrogen oxide emissions because the amount of excess air can be reduced below that used in coal or oil-fired stations, and because two-stage combustion of the fuel gas can reduce greatly the formation of the nitrogen oxides.

b. Higher levels of sulfur cleanup - potentially as high as 96%.

c. The process directly produces elemental sulfur, which can be stockpiled cheaply without danger of pollution. Not all of the stack gas treatment processes now under development allow recovery of elemental sulfur.

d. The fuel gas sulfur content is dictated by in situ reactions in the gasifier. Therefore, the sulfur recovery section of the plant need not be highly efficient.

A limitation of the CO_2 Acceptor Process is that its use would be restricted to stations which have adequate land available for installation of the fuel gas plant.

Basis for Evaluation

For evaluation of the cost of low-sulfur fuel gas, Case I was chosen. The product gas, at 1300°F and 206 psig, is passed through expanders which reduce the pressure to 25.7 psia. The expanded gas is delivered at 665°F to the burners at the boiler. The expanders drive compressors for the process air and produce an additional 170 MW of power.

The cost of the expander-compressor-generator system was estimated at $13.2 MM which, added to the "base" plant cost of $68 MM (page 50), gives the installed plant cost as $81.2 MM. Direct operating cost was increased over that shown in Table 20 to allow for increased maintenance.

Economics

The cost evaluation is summarized in Table I in the Summary. The fuel gas cost is shown on the bases both of the heat of combustion of the gas and of the heat of combustion plus the sensible heat content. The latter cost is more realistic since the sensible heat represents available heat to the boiler.

The impact of coal cost on fuel gas cost is shown in Figure 1.

From the standpoint of sulfur emissions, the CO_2 Acceptor Process converts coal having 4.3% sulfur content to a fuel gas which has a sulfur content equivalent to that in 0.3% sulfur content fuel oil, a material which is not now available in large quantity at any price.

In areas which impose severe restrictions of sulfur emissions, the costs shown in Figure 1 indicate that fuel gas from the CO_2 Acceptor Process would be competitive with low-sulfur imported fuel oil, with the added advantage of the lower NO_x emissions. Further, production of power from coal assures continuity of supply and reasonably predictable costs in contrast with politically-caused uncertainties involved with use of foreign oil.

Low-sulfur fuel oil now is being delivered to barges at East Coast ports at costs between 70-85¢/MM Btu, depending upon sulfur content and heating value.(12) Residual fuel oil with a sulfur level equivalent to that potentially possible from the processes considered here (ca. 0.3 wt % sulfur) is not available today. It would be necessary to use low-sulfur distillate fuels, at least in part, to meet these sulfur specifications.

IV. EQUIPMENT AND PROCEDURE

A. Description of the Continuous Unit

1. Process Piping and Equipment

A simplified flow diagram for the continuous gasification unit is shown in Figure 11. A more detailed process and instrumentation flow diagram is shown in Figure 12. These figures show the unit as arranged for "integrated" operation where the acceptor is recirculated continuously between the gasifier and regenerator.

Figure 13 shows a simplified flow diagram of the unit as arranged for single-vessel operation where the reactor (vessel D-1 in Figure 12) is used either as a preoxidizer or as a gasifier. The regenerator is not used since acceptor circulation is not required.

The following description applies to operation of the integrated system where the feedstock to the gasifier is a precarbonized char. In future operation of the integrated system, the char feedstock will be produced in a separate operation by partial gasification of preoxidized coal in the single-vessel system shown in Figure 13.

For single-vessel operation, the equipment and procedure in general are the same as described here except that all processing steps involving the acceptor and regenerator are eliminated. This mode of operation is described separately in Section IV-D, page 97.

a. Process Vessels

The continuous unit was designed on the basis of 2-10 lb/hr of char feed to the gasifier. The corresponding acceptor circulation rates required to maintain heat balance were in the range of 5-20 lb/hr (raw acceptor basis). These feed rates and the anticipated kinetics for the steam-carbon reaction and acceptor calcining reaction dictated the sizes of the process vessels.

The gasifier vessel D-1 shown in Figure 14 is 4" in diameter. The char bed height, fixed by an overflow weir is about 38", and the height of the acceptor bed directly below is about 7-1/2".

The regenerator vessel D-2 shown in Figure 15 is 3" in diameter. The bed height is maintained at 18" by an overflow weir. Details of the internals and fittings of both vessels are given in Figure 16.

Heat was supplied through the walls of the process vessels. The gasifier heater contained three separate circuits which corresponded to the bottom, middle and top zones of the reactor. The regenerator was equipped with four heater circuits spaced along the length of the reactor. All the gasifier heaters and the top zone regenerator heater were controlled by 270 volt Powerstats supplying 5.1 KW of electrical heat. The three lower heaters on the regenerator were controlled by 230 volt Powerstats supplying 3.8 KW.

The process vessels and their heaters were contained in pressure shells which were pressurized with CO_2, so that the pressure drops across the walls of the process vessels are zero. Figures 17 and 18 show the gasifier and regenerator pressure shells, respectively.

Johns-Manville Sil-O-Cel C-3 insulation was used to fill the annular space between the heaters and the wall of the regenerator pressure shell. The regenerator pressure shell wall temperature was held at 950°F. The annular space between the heaters and wall of the gasifier pressure shell was filled with Carborundum "Fiberfrax" insulation. The gasifier pressure shell wall temperature was held below 1100°F.

b. Solids Handling

Seven solids streams were handled in the continuous gasification unit:

(1) Char feed to the gasifier.
(2) Char product from the gasifier.
* (3) Fuel char feed to the regenerator.
(4) Recarbonated acceptor to the regenerator.
(5) Calcined acceptor from the regenerator to the gasifier.
(6) Recarbonated acceptor from the gasifier.
(7) Overhead ash material from the regenerator.

* Fuel char is ungasified product char from the gasifier which is burned in the regenerator.

The residual fuel char was stripped from the regenerator bed and carried out overhead in the regenerator outlet gas stream. The overhead solids were removed from the gas by an external cyclone G-1.

The acceptor entering the gasifier in Stream 5 recarbonated as it showered down through the gasifier char bed, and then segregated at the bottom forming a sharply defined interface with the char bed. This was stabilized by means of a 2-1/2" diameter boot on the bottom of the char bed.

Solids streams 1, 3, and 4, were conveyed in electrically heated pneumatic transfer lines.

The recarbonated acceptor and fuel char were introduced at the bottom of the regenerator bed through a single axial tube so that the sensible heat of the acceptor and the heat of reaction of the endothermic calcining reaction would provide a heat sink for the heat released by the combustion of fuel char.

Streams 2, 5, and 6 were conveyed by gravity flow in purged standlegs. These standlegs formed the seals which prevent cross-contamination between the H_2-bearing gasifier gas and O_2-bearing regenerator gas. The solids flow rates in Streams 2 and 5 were controlled to maintain the standlegs full of solids and the solids flow rate in Stream 6 was controlled to maintain a constant char-acceptor interface level. To protect the Teflon seats in the valves used to control the solids flow rates, water-jacketed coolers were installed in each of the standlegs. These coolers are specified in Figure 19. Sight glasses (Jerguson 12-T-20) were also installed in each standleg to allow visual confirmation of the amount and nature of solids flow.

The gasifier char overflow weir was placed inside a compartment which occupied 30% of the vessel cross-section to prevent contamination of the product char by the acceptor which entered the gasifier.

The calcined acceptor entered the top flange of the gasifier and passed through a preheater purged with N_2 which was introduced into the acceptor return line above the vessel. The preheater was a stainless steel tube, 7/8" OD x 3/4" ID x 10" long having nine baffles installed at 60° angles. In the preheater the acceptor was heated to about 1200°F before falling into the char bed.

Feed hoppers and withdrawal hoppers were provided so that the solids flow rates in Streams 1, 3, and 4 could be controlled independently of the inventory requirements of the fluidized beds and standlegs.

The gasifier char feedstock was supplied from one of two char feed hoppers, F-1A and F-1B, each having a capacity of about 100 pounds. Continuity of feed was maintained by feeding char alternately from the two feed hoppers.

Gasifier product char was withdrawn into a lockhopper F-2 at a rate dictated by the overflow in the weir. The capacity of the lockhopper was approximately ten pounds.

Recarbonated acceptor from the gasifier was withdrawn into one of two lockhoppers, F-6A and F-6B, at a rate dictated by the level of the gasifier char acceptor interface. The acceptor was removed from the off-stream lockhopper and charged back to one of two acceptor feed hoppers, F-4A and F-4B. Continuity of acceptor circulation was maintained by feeding acceptor to the regenerator alternately from the two feed hoppers. A sight glass was installed below the feed hoppers to show when the on-stream feed hopper was empty. The feed hoppers and withdrawal hoppers each have a capacity of six pounds of raw acceptor.

A lockhopper F-5 was installed on the acceptor return line between the regenerator and the gasifier to allow addition of acceptor to the gasifier. The capacity of the acceptor charge pot was 1.1 pounds of raw acceptor.

A sampling tee was inserted in the L-5 acceptor return line through which a sample of calcined acceptor could be withdrawn while the unit is operating.

c. Gas Flows

The gasifier inlet gas consisted of recycled product gas with H_2S, H_2, CO_2, and steam flows added to give the desired gas composition.

The metered dry gasifier fluidizing gas was passed through the steam generator which is shown in Figure 20. The water temperature in the steam generator was controlled (±0.5°F) to give the desired steam partial pressure. Calibrations showed that at any given water temperature, in the range of 295 to 370°F, the steam partial pressure was equal to the equilibrium vapor pressure of water. The steam generator feed water was passed through a Betz Laboratory Demineralizer.

The fluidizing gas passed through an electrically heated line from the steam generator to the top of the gasifier. The wall temperature in the line was kept above 400°F to prevent condensation of the steam. The gas inside the gasifier vessel passed through a helically-wound preheat coil and an axial dip tube which extended into a cone at the bottom of the vessel. The gas passed downward, reversed direction, and fluidized the bed.

The gasifier outlet gas passed through one of two parallel filters where solids carried over from the gasifier bed were removed using F-porosity Micro-Metallic filters. Both of the filters and the upstream piping were heated electrically to prevent condensation of the unreacted steam in the outlet gas. Separate heating circuits were provided for each of the filters so that they could be cooled before being removed for cleaning and heated rapidly after being reinstalled.

The unreacted steam was condensed in the C-1 cooler. The condensed steam was collected in the condensate receiver F-8 which was mounted below C-1. A liquid level controller actuated a control valve C-7 to drain condensate and maintain a constant liquid level in the receiver.

The gas was further cooled in the C-4 cooler using water supplied at 40°F from a water chilling unit. Additional condensate was disengaged in knock-out pot F-9 and collected in the sight glass below F-9.

The dry solids-free product gas was vented to atmospheric pressure through the gasifier back pressure control valve PCV-1.

The regenerator inlet gas consisted of recycled regenerator product gas with air, CO_2, and N_2 added to give the desired concentrations of CO_2 and CO or air in the product gas.

About two-thirds of the inlet gas entered the top of the regenerator and passed through a helically-wound preheat coil and an axial dip tube which extended into the cone at the bottom of the vessel. The gas passed downward, reversed direction, and fluidized the bed.

The remainder of the regenerator inlet gas consisted of the acceptor and fuel char carrier gas which entered the top of the regenerator and passed through the axial solids inlet tube to the bottom of the vessel.

The regenerator outlet gas passed first through an external cyclone G-1, shown in Figure 21, which removed about 98% of the fuel char residue. The remainder of the overhead solids were removed in one of two parallel filters which used F-porosity Micro-Metallic filter elements.

The regenerator gas was cooled in a cooler, C-6, which was supplied with water at 40°F from the water chilling unit. The steam formed by combustion of the hydrogen in the fuel char was condensed and collected in a sight glass below the cooler. The gas coolers, C-1, C-4, and C-6, are specified in Figure 22.

The dry, solids-free regenerator product gas was vented to atmospheric pressure through the back pressure control valve DPCV-1.

Small metered flows of N_2 were used to purge the pressure taps at the tops of both vessels as well as the L-2, L-6, and L-5 standlegs, the acceptor return line, and the sheath which contained the leadwire for the char-acceptor interface probe. A small flow of N_2 was added to the on-stream gasifier char feed hopper and a small flow of N_2 added to the fuel char feed hopper to compensate for the solids fed. All of the purge flows were as small as possible relative to the fluidizing gas flows.

The H_2, CO_2, and N_2 used as process gas, pressure shell balance gas, and lockhopper pressurizing gas were supplied from cylinders. Regulators downstream of the cylinder manifolds fixed the supply pressures at 300 psig. Air was supplied from a compressor. Water and oil were removed by a knock-out drum followed by a bed of silica gel.

Identical single-stage diaphragm compressors (Pressure Products Industries, Model 1073) were used to recycle gasifier and regenerator product gas.

d. Miscellaneous

Wherever possible, connections to the process piping were high pressure coned fittings (Autoclave Engineers or Pressure Products) which were used because of their ruggedness, ease of disassembly, and freedom from leaks.

Special adapter fittings were required to connect the solids outlet tubes (3/8 x .035" wall) to the standleg piping (9/16" OD x 5/16" ID). The solids outlet tubes from the vessels slipped inside the adapter fittings and were sealed using a lava sealant and packing gland. A standard high-pressure connection was made between the adapter fitting and the standleg.

In order to minimize the pressure drop across the gas outlet piping, 3/8 x .028" wall tubing was used downstream of C-1 and C-6 coolers in the gasifier and regenerator, respectively.

The steam condensate contained dissolved NH_3, SO_2, H_2S, and CO_2. Type 304 stainless steel gave excellent service in piping and equipment handling process gases at temperatures from ambient to 400°F. Previous experience had shown that condensate and wet process gases are extremely corrosive to copper, brass and plain carbon steel. These materials were not used, even in instrumentation piping.

Conax packing glands and lava sealants were used where the heater leadwires, solids outlet tubes, and the vessel skin thermocouples passed through the pressure shells.

2. Instrumentation and Control

a. Pressure Measurement and Control

The system pressure was controlled at the top of the gasifier vessel using a Foxboro pressure cell and recorder-controller which operated the back pressure control valve PCV-1. The pressure at the top of the regenerator was maintained at a fixed differential (usually in the range of zero to five inches H_2O) from that at the top of the gasifier using a Foxboro d/P cell, and a recorder-controller which operated the regenerator back pressure control valve DPCV-1

The discharge pressures of the two recycle compressors were controlled using Foxboro indicator controllers which operated control valves PCV-2 and PCV-3, which vented to the suction side of the compressors.

Identical balance gas systems were used to maintain a zero pressure differential across the walls of both process vessels. CO_2 was supplied to the pressure shell at 300 psig. Barton d/P cells and recorder-controllers operated control valves DPCV-11A and 11B and DPCV-13A and 13B which either loaded the shell or vented CO_2 to the atmosphere to maintain a zero pressure differential between the pressure shell and the process vessel.

The unit was adequately instrumented to monitor pressure drops across the various transfer lines, standlegs, and fluidized beds. These pressure drops were measured with respect to the pressures at the tops of the gasifier or regenerator vessels, and were continuously recorded.

The pressure drops across the solids filters in both the gasifier and regenerator outlet gas piping were continuously recorded. The filters were switched when the pressure drop across the on-stream filter reached 75" of H_2O.

Panel mounted pressure gauges were used to indicate the pressures at the top of both vessels, the pressure shell pressures, the compressor discharge pressures, the char and fuel char feed hopper pressures, and the pressures immediately upstream of the back pressure control valves. Locally mounted pressure gauges were used to indicate the pressures in the steam generator and the various lockhoppers.

b. Temperature Measurement and Control

The bed temperatures in both the gasifier and regenerator were measured using chromel-alumel thermocouples contained in wells which were immersed in the fluidized beds. Two thermocouples were used in the gasifier thermowell and one in the regenerator.

The bed temperatures were controlled manually by adjusting the Powerstats which controlled the electrical heat input from the heaters.

In each vessel, the wall temperatures corresponding to each of the heating zones were measured using thermocouples which were inserted through the side of the pressure shell and through slots ground in the heater ceramics.

The water temperature in the steam generator was monitored using an Iron-Constantan thermocouple contained in an axial thermowell. The thermocouple signal was recorded using a Foxboro temperature recorder-controller. A high or low temperature activated relays which supplied power to the steam generator bayonet heater from the appropriate one of two 7.5 KW Powerstats which were set by trial to supply slightly more and slightly less than the required steady-state input.

A total of 50 temperatures were monitored of which 31 were continuously recorded. The other temperatures included the wall temperatures of the pressure shells, the transfer lines, the inlet and outlet gas piping, the steam generator, the gasifier solids filters, the regenerator cyclone, and the ash receiver. Thermocouples inserted in the sight glasses in the various standlegs monitored the temperatures of the solids flowing from the water-jacketed coolers.

c. Control of Solids Flow Rates

The solids flow rates were controlled using rotary feeders which consisted of a tapered Teflon plug rotating inside a stainless steel body. Drawings of the feeders used in the continuous unit are shown in Figures 23, 24, and 25. The number, diameter, and depth of the pockets were chosen on the basis of the bulk density of the solid and the desired range of solids flow rates. Variation of the solids flow rates over the desired range was accomplished by means of a Graham variable speed drive. Slots were milled in the feeder body around the inlet and outlet ports to allow the pockets to begin to fill and discharge before reaching top and bottom dead center. Since Teflon was used, no lubrication was required between the feeder body and rotor.

The solids feeders supplied an essentially continuous flow of solids over the desired range of feed rates. The solids feed rate was linear with rotor rpm up to the maximum speed of 12 rpm.

A rotary feeder could not be used to feed calcined acceptor. Due to lack of physical strength, this material would be severely attrited by the shearing action of the rotor turning in the stainless steel body. The flow of calcined acceptor was controlled using a Hoke ball valve having a 5/16" bore.

d. Solids Level Control

To control the level of the gasifier char-acceptor interface, a curved type 310 stainless steel plate, 2" long x 1-1/2" wide x 1/16" thick, was positioned as shown in Figure 14. The plate was supported by a 1/16" diameter 310 stainless steel leadwire which entered the top of the gasifier vessel. The leadwire was insulated from the char bed by a 1/4" OD x .028" wall type 310 stainless steel sheath, which was purged with N_2. Alumina beads, .068" ID x .170" long, were spaced at 2" intervals along the leadwire to prevent its shorting against the sheath.

The dc circuit shown in Figure 25 was connected across the plate and the vessel walls and internals. The circuit was designed to give a 50 millivolt signal when the plate was covered completely by the electrically conducting char, and essentially zero signal when the plate was covered by the non-conducting acceptor. The L-6 valve, which fixed the rate of acceptor withdrawal from the acceptor layer, was operated at one of three settings to control the interface level. The medium setting was chosen to correspond to the nominal acceptor circulation rate. A higher setting was used if the interface level was increasing and a lower setting was used if the interface level was decreasing. The best control occurred when the lower setting was zero.

Two rotax switches were attached to the pen arm of the controller, which recorded the circuit signal. They were activated at pen positions corresponding to 15 and 35 millivolt signals. Upon being activated, a switch would open and close (or close and open) the appropriate two of three solenoid valves which supplied air at a preset pressure in the 3 to 15 psig range to the Conoflow Cylinder Operator used to position the control shaft of the valve.

To detect the level of calcined acceptor in the standleg above the L-5 valve, a thermocouple was installed in the adapter section between the bottom of the regenerator pressure shell and the water-jacketed standleg. The thermocouple extended 1/16" into the 5/16" bore of the adapter. When hot acceptor, which had been discharged from the regenerator, immersed the thermocouple; a temperature increase occurred. The temperature controller which read out the thermocouple signal opened the L-5 valve, allowed acceptor to flow into the gasifier and thereby lowered the acceptor level in the standleg. When the thermocouple was uncovered, the temperature fell off and the temperature controller reduced the L-5 valve opening to stop the flow of acceptor until the thermocouple again was immersed in the hot acceptor which overflowed the regenerator weir.

The temperature controller opened and closed (or closed and opened) two solenoid valves which supplied air at preset pressures, in the 3 to 15 psig range, to the Conoflow Cylinder Operator used to position the L-5 valve. When the temperature was below the control index, the L-5 valve opening was set so that purge gas would flow through the valve but acceptor would not. When temperature was above the index, the L-5 valve was opened to allow acceptor to flow through the valve at a rate somewhat higher than the nominal acceptor circulation rate.

The operation of the L-5 valve was monitored continuously by recording the air pressure supply to the Conoflow Cylinder Operator.

Since acceptor in this standleg acts as a seal between the gasifier and regenerator vessels, an auxiliary level control system was used to over-ride the normal control system and close L-5 valve should the level drop to the level of the sight glass. A photocell circuit was used to detect the interface level. A rotax switch was attached to the pen arm of the controller which recorded the output of the circuit. When the circuit signal reached 35 millivolts, the switch activated closing the open solenoid valve in the normal control system and opened a third solenoid valve which vented the air supply to the Conoflow Cylinder Operator and closed the L-5 valve.

e. Gas Flow Measurements

The inlet gas flows were metered using calibrated rotameters which were calibrated with either a Rockwell dry gas meter or a wet test meter depending on the range of the rotameter. The gas meters were calibrated by passing known weights of CO_2 through them.

Rotameters using recycle gas were calibrated using two gases having different molecular weights. The flow rates were correlated with the rotameter float heights using the following general equation:

$$Q = k(M_r)^{-n} \qquad (9)$$

Where:
Q = flow rate, SCFH.
k = constant for a given float height.
M_r = molecular weight of the recycle gas.
n = constant for a given rotameter at a given supply pressure.

The constant n varied between .455 and .500 for five rotameters calibrated at supply pressures of 400 and 250 psig.

A plot of k versus float height was made for each of the rotameters at both supply pressures.

The dry product gas flow rates from both vessels were measured using calibrated Rockwell dry gas meters located downstream of the back pressure control valves.

f. Gas Analyzers

The product gas streams from both vessels were analyzed using a gas chromatograph. During a run, low levels of CO, CO_2, and SO_2 were continuously monitored with infrared analyzers. High levels of CO_2 were monitored with a thermal conductivity cell with reference to nitrogen. A paramagnetic susceptibility analyzer was used to measure the concentration of oxygen. The outputs of the CO, CO_2, SO_2, O_2 analyzers and of the thermal conductivity cell all were continuously recorded. Analyses were made on the gas chromatograph when the above analyzers indicated that the gas composition was steady.

All of the gas analysis instruments were connected to three-way valves so that offgas from either the regenerator or gasifier could be fed to each. For integrated operation the chromatograph analyzed both the regenerator and the gasifier offgases. The infrared CO_2 analyzer monitored the gasifier, and the remaining instruments checked the regenerator offgas. In addition, gas samples were taken into a stainless steel cylinder for intensive chromatographic analysis by the analytical laboratory.

3. Safety Features

The unit was equipped to give an audible alarm whenever a potentially dangerous upset occurred. A light indicating the source of the upset accompanied the alarm. In some situations, relays were activated which closed the emergency solenoid valve to isolate the gasifier and regenerator vessels and shut off appropriate solids feeders.

The possibility of an explosion exists if cross-contamination of the gasifier and regenerator process gases occurred. During normal operations, cross-contamination was prevented by maintaining the N_2 purged standleg above the L-5 valve full of acceptor and by controlling the pressure differential between the tops of the gasifier and regenerator vessels such that the pressure drop across the L-5 valve was zero. The presence of solids in the L-5 standleg was guaranteed by the photocell circuit in the standleg level control system which closed the L-5 valve when an acceptor interface appeared in the sight glass. If an upset occurred such that the pressure differential between the vessels deviated ± 10" H_2O from the set valve or the pressure drop across the L-5 valve deviated ± 5" H_2O from zero, an alarm sounded and the emergency solenoid valve closed, thereby isolating the gasifier and regenerator vessels.

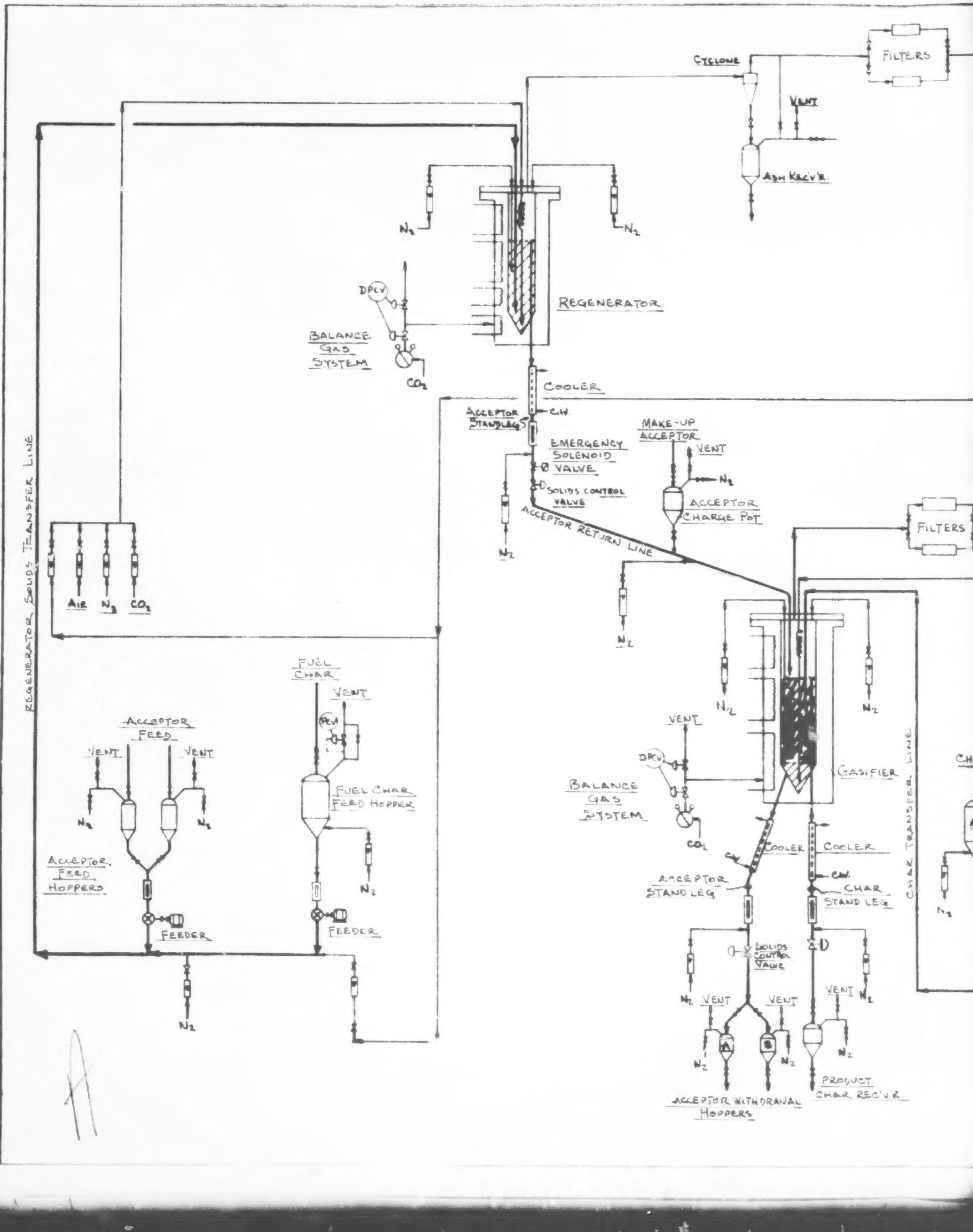
Recycle gas, which was used to transport the gasifier char feedstock and the acceptor to the regenerator, was a major constituent of the fluidizing gases of both vessels, and was vital for operation of the unit. Pressure switches were installed to sound an alarm, close the emergency solenoid valve, and shut off the L-4 feeder if the discharge pressure of either recycle gas compressor fell below a preset valve. The L-1 feeder was also shut off if the gasifier recycle compressor discharge pressure was low.

The supply of process air and feed of fuel char and acceptor, which were vital to the control of the regenerator temperature, were monitored using alarms. High and low ΔP alarms were installed to detect plugging or starvation in either the acceptor or fuel char transfer lines. Other alarms warned of low process air pressure and high oxygen concentration in the regenerator product gas.

The unit was equipped with a high ΔP alarm to detect plugging in the gasifier inlet piping. Another alarm sounded in the event of malfunctioning of the gasifier back pressure valve or of plugging in the gasifier outlet piping, as indicated by a high pressure at the top of the vessel. Other alarms warned of malfunction of the liquid discharge valve below the condensate receiver, as indicated by a high or low condensate level.

Failure of the balance gas system to control pressure differential across the walls of either process vessel could have resulted in collapse or rupture of the vessel. An alarm was provided to indicate any deviation in the pressure differential between either process vessel and its pressure shell greater than 25" of water and another alarm was used to indicate a low pressure in the CO_2 cylinder manifold which supplied the balance gas.

Nitrogen was available to supply pressure to the instruments if the supply of instrument air were lost, as shown by an alarm indicating low instrument air supply pressure. Cooling water was available from a cooling tower if the city water supply were cut off, as indicated by a low pressure alarm.


CYCLONE
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ASH RECV'R.
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N2
DPCV
BALANCE GAS SYSTEM
CO2
REGENERATOR
COOLER
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ACCEPTOR STANDLEG
EMERGENCY SOLENOID VALVE
SOLIDS CONTROL VALVE
MAKE-UP ACCEPTOR
VENT
N2
ACCEPTOR CHARGE POT
ACCEPTOR RETURN LINE
N2
FILTERS
N2
REGENERATOR SOLIDS TRANSFER LINE
AIR
N2
CO2
N2
N2
N2
FUEL CHAR
VENT
PCV
ACCEPTOR FEED
VENT
VENT
N2
N2
FUEL CHAR FEED HOPPER
ACCEPTOR FEED HOPPERS
N2
FEEDER
FEEDER
N2
VENT
DPCV
BALANCE GAS SYSTEM
CO2
GASIFIER
CHAR TRANSFER LINE
C.W.
COOLER
COOLER
C.W.
ACCEPTOR STAND LEG
CHAR STAND LEG
SOLIDS CONTROL VALVE
N2
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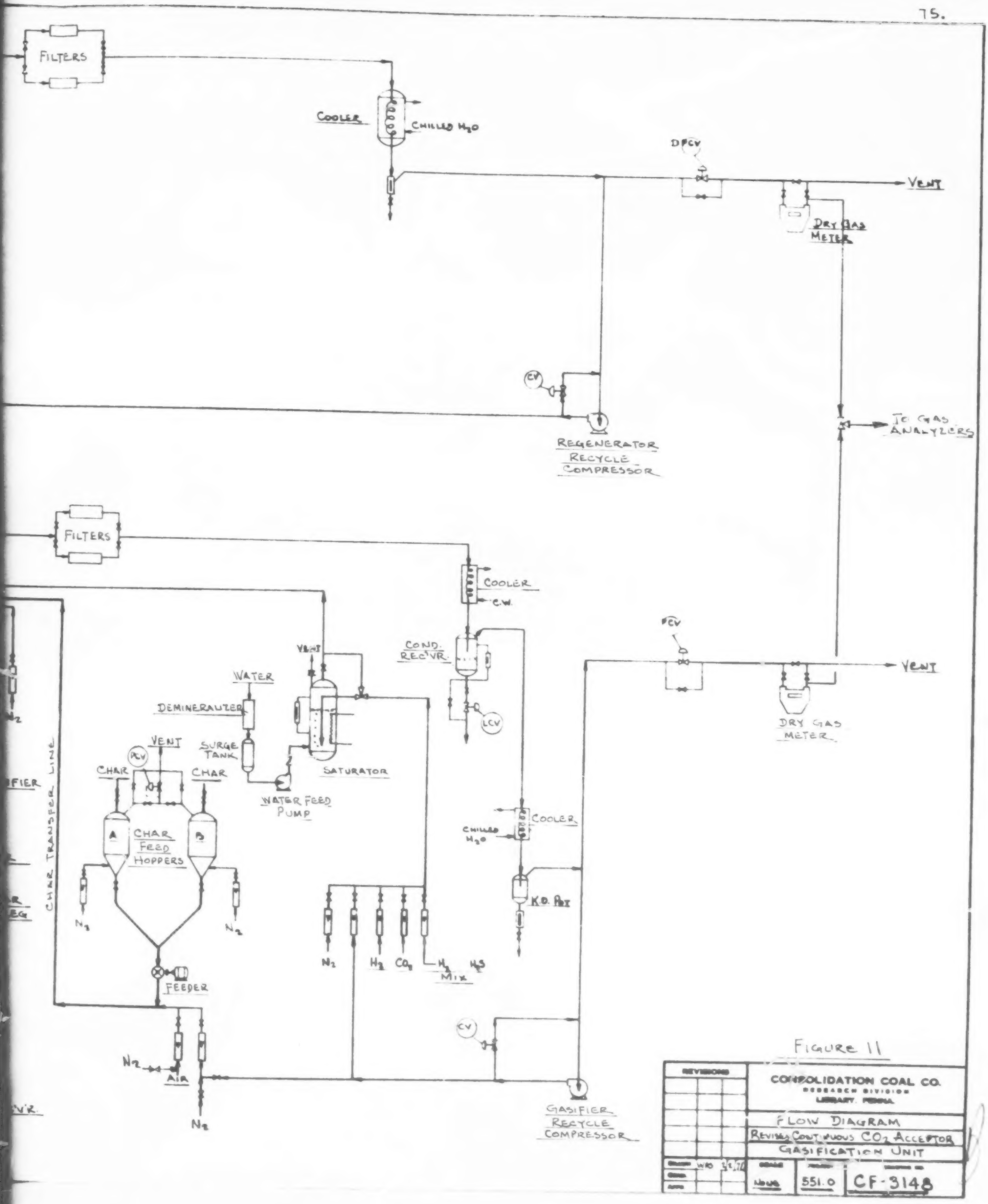
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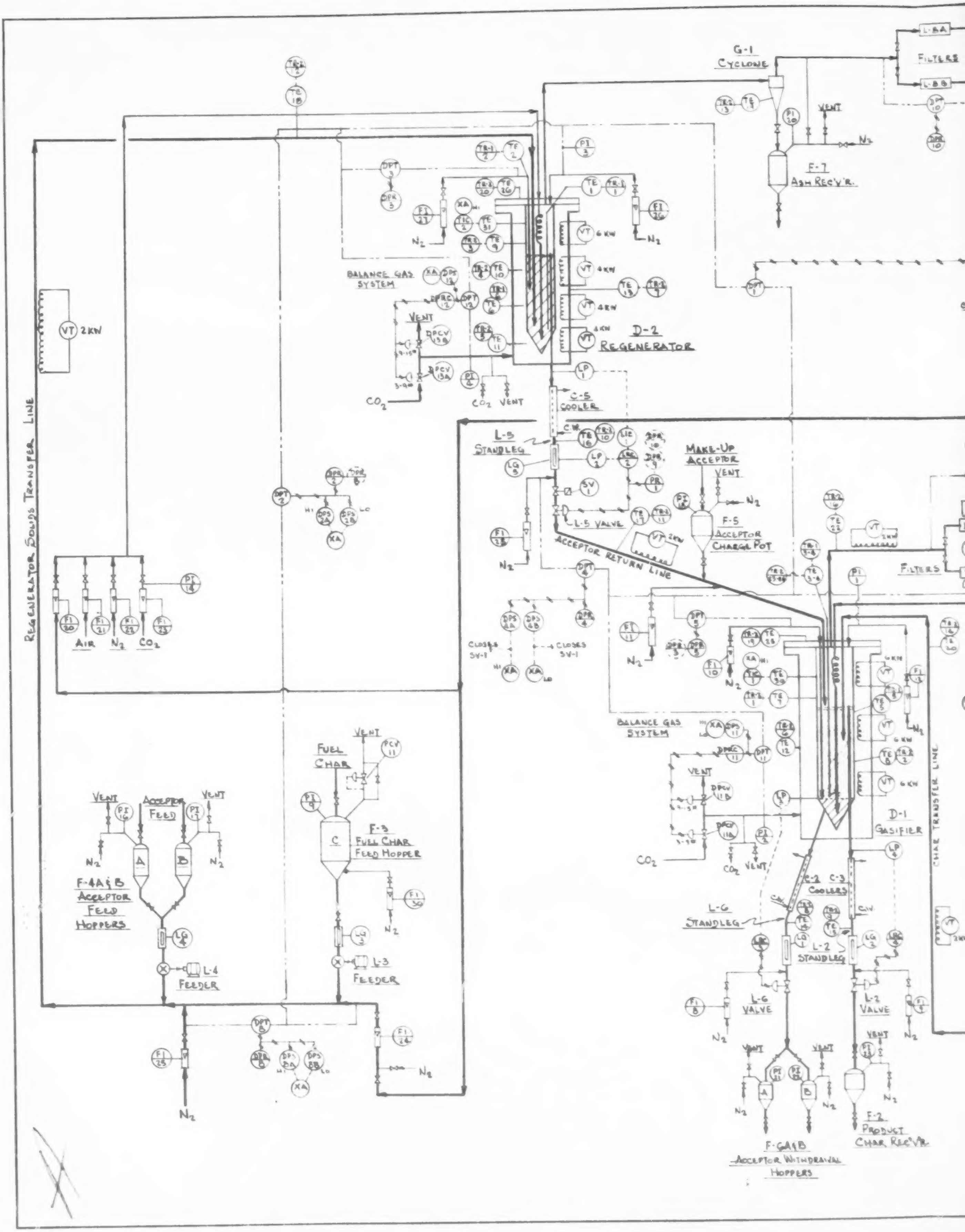
FILTERS
COOLER
C.W.
COND. REC'VR.
LCV
FCV
VENT
DRY GAS METER
VENT
WATER
DEMINERALIZER
SURGE TANK
SATURATOR
WATER FEED PUMP
PCV
VENT
CHAR
CHAR
A
B
CHAR FEED HOPPERS
N_2
N_2
FEEDER
CHAR TRANSFER LINE
N_2
COOLER
CHILLED H_2O
K.O. POT
N_2
H_2
CO_2
H_2 H_2S MIX
CV
N_2
AIR
N_2
GASIFIER RECYCLE COMPRESSOR
REVISIONS
DRAWN WKO 3/2/70
CHKD.
APPD.
SCALE
NONE
PROJECT
551.0
DRAWING NO.

FIGURE 11

CONSOLIDATION COAL CO.
RESEARCH DIVISION
LIBRARY, PENNA.

FLOW DIAGRAM
REVISED CONTINUOUS CO_2 ACCEPTOR
GASIFICATION UNIT

CF-3148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FILTERS
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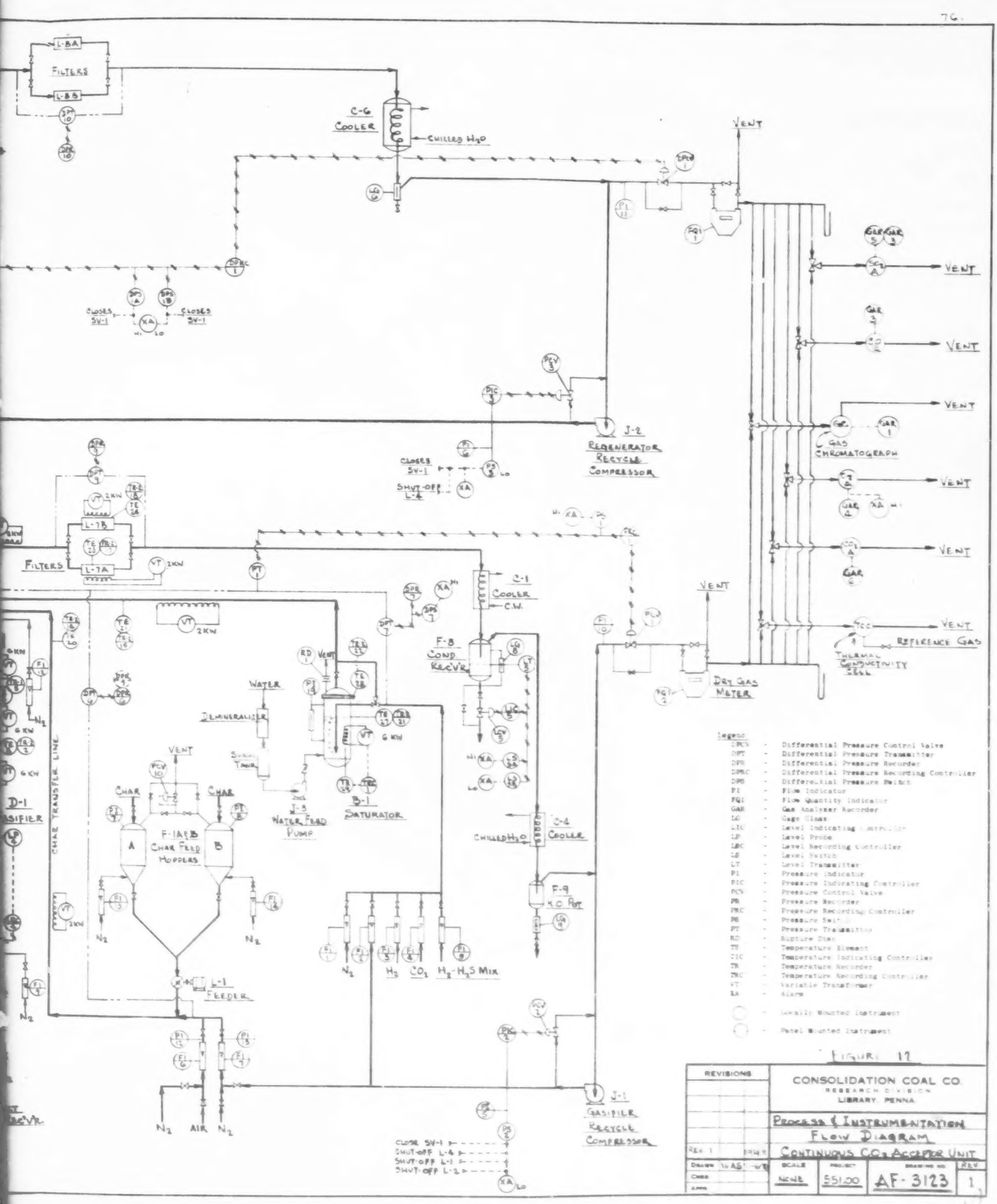
L-8B
VENT
N2
F-7 ASH REC'V'R.
D-2 REGENERATOR
VT 6 KW
VT 4 KW
VT 2 KW
BALANCE GAS SYSTEM
VENT
CO2
CO2 VENT
C-5 COOLER
C.W.
L-5 STANDLEG
L-5 VALVE
MAKE-UP ACCEPTOR
F-5 ACCEPTOR CHARGE POT
ACCEPTOR RETURN LINE
REGENERATOR SOLIDS TRANSFER LINE
AIR
N2
CO2
CLOSES SV-1
FILTERS
BALANCE GAS SYSTEM
D-1 GASIFIER
CHAR TRANSFER LINE
C-2 C-3 COOLERS
L-6 STANDLEG
L-2 STANDLEG
L-6 VALVE
L-2 VALVE
F-6A & B ACCEPTOR WITHDRAWAL HOPPERS
F-2 PRODUCT CHAR REC'V'R.
FUEL CHAR
F-3 FUEL CHAR FEED HOPPER
L-3 FEEDER
ACCEPTOR FEED
F-4A & B ACCEPTOR FEED HOPPERS
L-4 FEEDER
A
B
C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L-8B
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Closes SV-1
XA
Gas Chromatograph
Vent
J-2 Regenerator Recycle Compressor
Closes SV-1
Shut-off L-4
Filters
L-7B
L-7A
2KW
C-1 Cooler
C.W.
F-8 Cond Recv'r
Dry Gas Meter
Thermal Conductivity Cell
Reference Gas
D-1 Gasifier
Char Transfer Line
Water
Demineralizer
Surge Tank
J-3 Water Feed Pump
B-1 Saturator
6 KW
Char
Char
Vent
F-1A&B Char Feed Hoppers
A
B
N2
N2
H2
CO2
H2-H2S Mix
C-4 Cooler
Chilled H2O
F-9 K.O. Pot
L-1 Feeder
N2
Air
N2
J-1 Gasifier Recycle Compressor
Close SV-1
Shut-off L-4
Shut-off L-1
Shut-off L-2

Legend

DPCV	-	Differential Pressure Control Valve
DPT	-	Differential Pressure Transmitter
DPR	-	Differential Pressure Recorder
DPRC	-	Differential Pressure Recording Controller
DPS	-	Differential Pressure Switch
FI	-	Flow Indicator
FQI	-	Flow Quantity Indicator
GAR	-	Gas Analyzer Recorder
LG	-	Gage Glass
LIC	-	Level Indicating Controller
LP	-	Level Probe
LRC	-	Level Recording Controller
LS	-	Level Switch
LT	-	Level Transmitter
PI	-	Pressure Indicator
PIC	-	Pressure Indicating Controller
PCV	-	Pressure Control Valve
PR	-	Pressure Recorder
PRC	-	Pressure Recording Controller
PS	-	Pressure Switch
PT	-	Pressure Transmitter
RD	-	Rupture Disc
TE	-	Temperature Element
TIC	-	Temperature Indicating Controller
TR	-	Temperature Recorder
TRC	-	Temperature Recording Controller
VT	-	Variable Transformer
XA	-	Alarm
○	-	Locally Mounted Instrument
⊖	-	Panel Mounted Instrument

Figure 17

REVISIONS

CONSOLIDATION COAL CO.
RESEARCH DIVISION
LIBRARY, PENNA.

Process & Instrumentation
Flow Diagram
Continuous CO2 Acceptor Unit

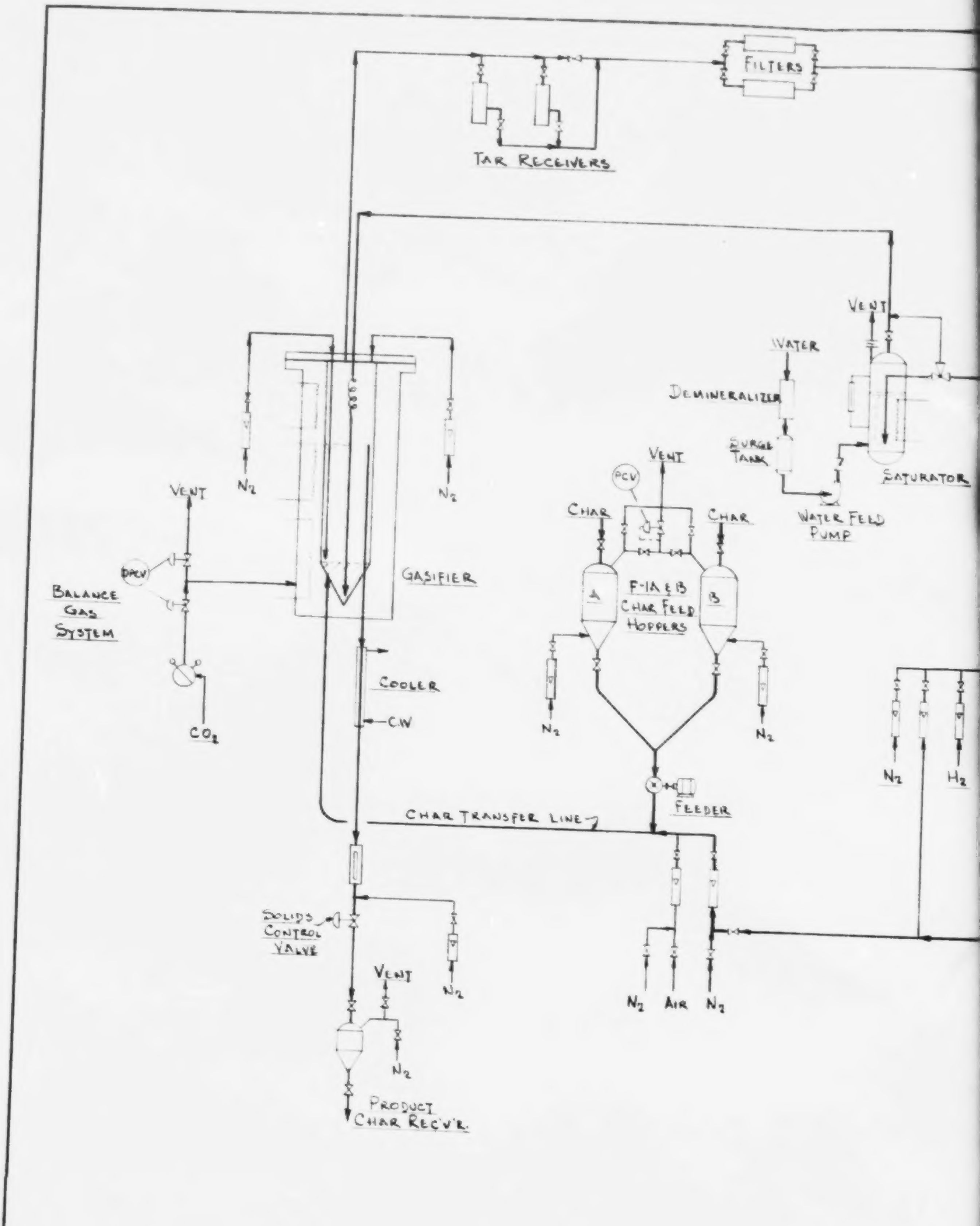
Rev. 1

Drawn W.A.S.

Chkd

Appd

Scale	Project	Drawing No.	Rev
None	551.00	AF-3123	1


TAR RECEIVERS
FILTERS
VENT
WATER
DEMINERALIZER
SURGE TANK
SATURATOR
WATER FEED PUMP
VENT
PCV
CHAR
CHAR
A
B
F-1A & B CHAR FEED HOPPERS
N$_2$
N$_2$
N$_2$
N$_2$
VENT
DPCV
BALANCE GAS SYSTEM
GASIFIER
CO$_2$
COOLER
C.W
FEEDER
CHAR TRANSFER LINE
N$_2$
H$_2$
SOLIDS CONTROL VALVE
VENT
N$_2$
N$_2$
AIR
N$_2$
N$_2$
PRODUCT CHAR REC'V'R.

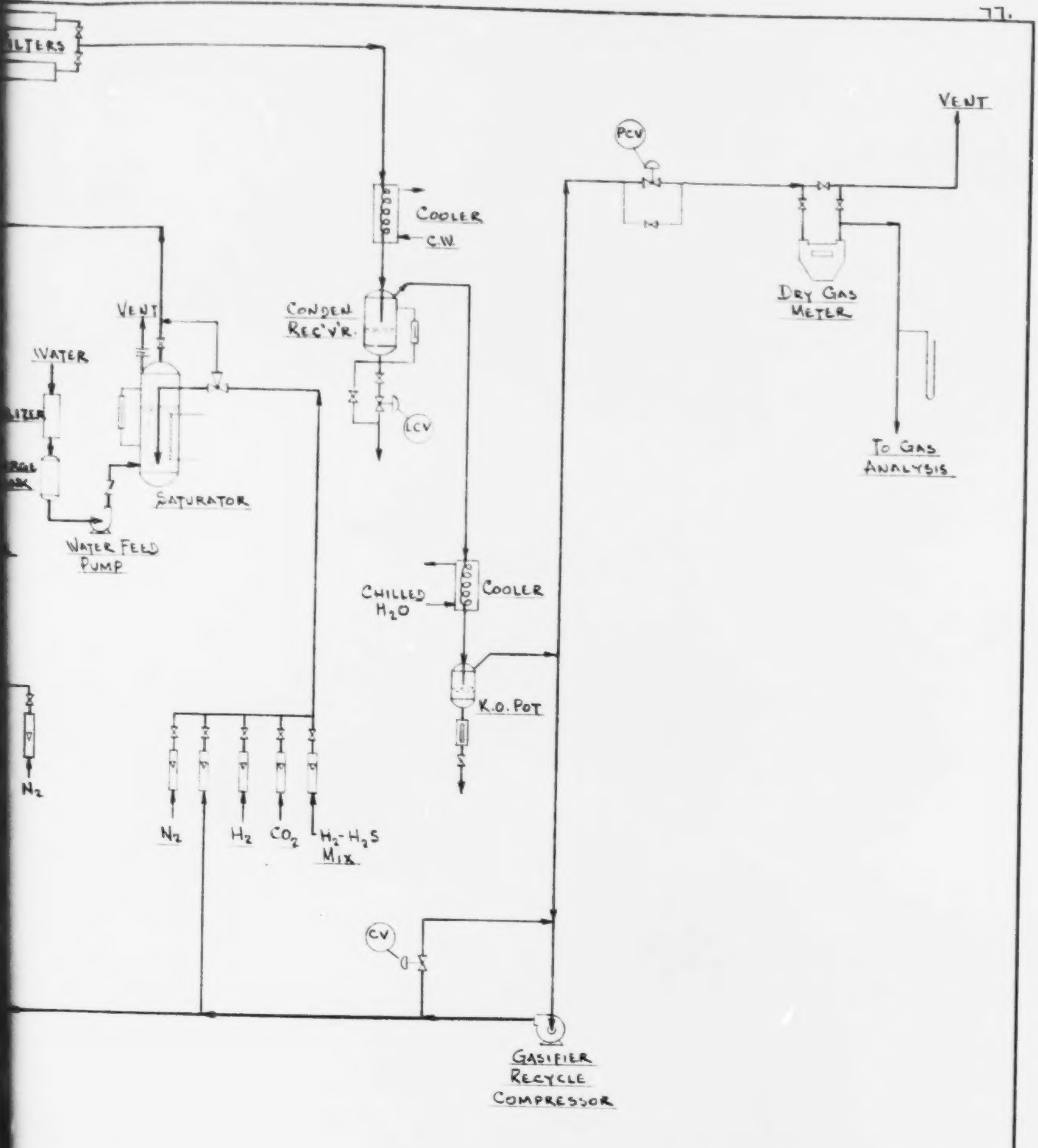
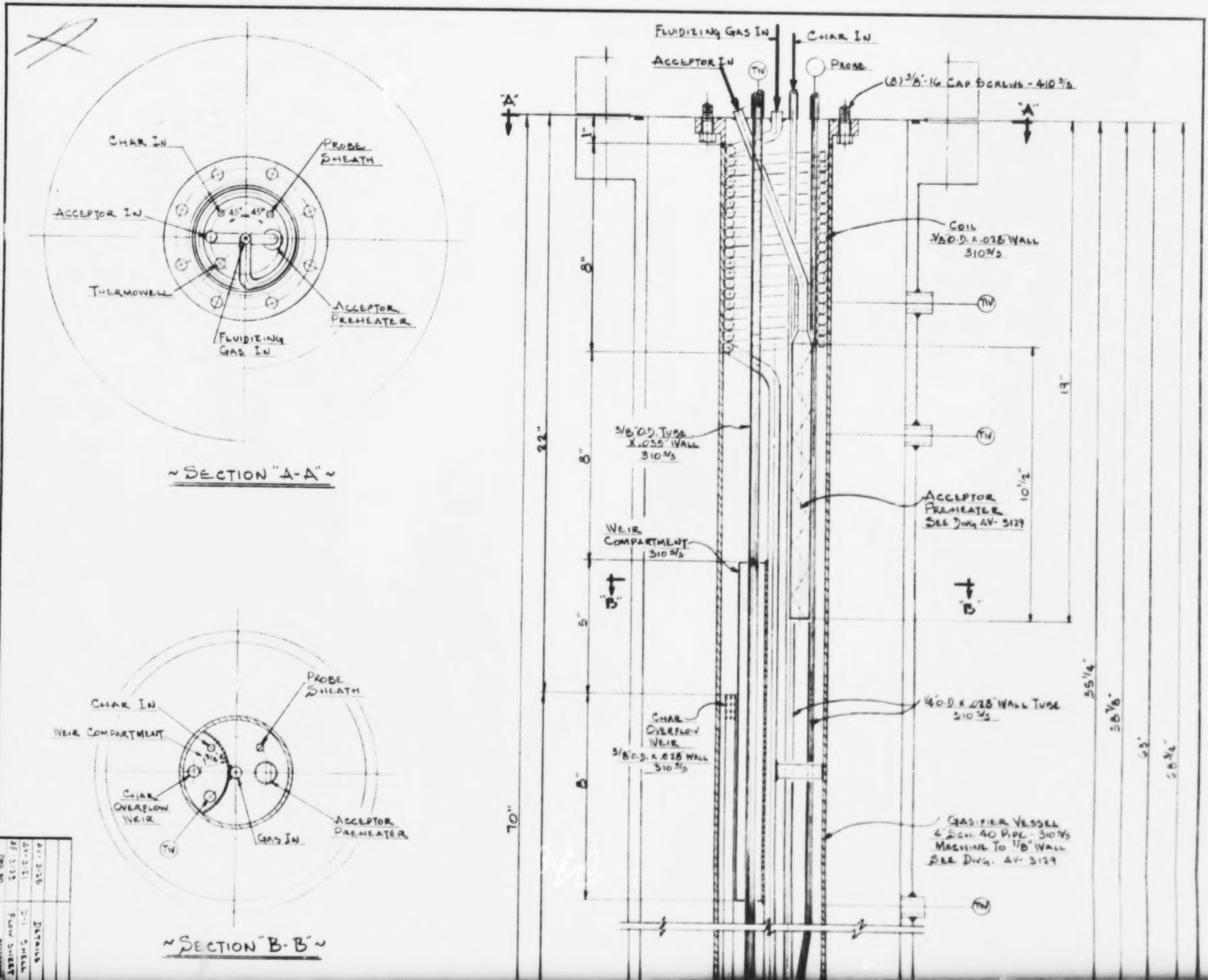

LTERS
VENT
PCV
COOLER
C.W.
DRY GAS
METER
CONDEN.
REC'V'R.
VENT
WATER
IZER
RGE
NK
SATURATOR
LCV
TO GAS
ANALYSIS
WATER FEED
PUMP
CHILLED
H_2O
COOLER
K.O. POT
N_2
N_2
H_2
CO_2
H_2-H_2S
MIX
CV
GASIFIER
RECYCLE
COMPRESSOR

FIGURE 13

REVISIONS			CONSOLIDATION COAL CO. RESEARCH & DEVELOPMENT DIVISION LIBRARY, PENNA.		
			FLOW DIAGRAM CONTINUOUS GASIFICATION - PREOXIDATION UNIT		
DRAWN	WAS	[illegible]	SCALE	PROJECT	DRAWING NO.
CHKD.			NONE	552.00	BF-3224
APPR.					

B


FLUIDIZING GAS IN
CHAR IN
ACCEPTOR IN
TW
PROBE
(8) 3/8"-16 CAP SCREWS - 410 S/S
"A"
"A"
1"
8"
8"
5"
8"
22"
70"
19"
10 1/2"
55 1/4"
58 7/8"
63"
68 3/4"
COIL
1/4"O.D. x .028" WALL
310 S/S
TW
TW
TW
3/8"O.D. TUBE
X .035" WALL
310 S/S
ACCEPTOR
PREHEATER
SEE DWG AV-3129
WEIR
COMPARTMENT
310 S/S
"B"
"B"
CHAR
OVERFLOW
WEIR
3/8"O.D. x .028 WALL
310 S/S
1/4"O.D. x .028" WALL TUBE
310 S/S
GASIFIER VESSEL
4" SCH. 40 PIPE - 310 S/S
MACHINE TO 1/8" WALL
SEE DWG. AV-3129
CHAR IN
PROBE
SHEATH
ACCEPTOR IN
45°
45°
THERMOWELL
ACCEPTOR
PREHEATER
FLUIDIZING
GAS IN
~SECTION "A-A"~
PROBE
SHEATH
CHAR IN
WEIR COMPARTMENT
3/16"
CHAR
OVERFLOW
WEIR
TW
GAS IN
ACCEPTOR
PREHEATER
~SECTION "B-B"~
AV-3125 DETAILS
AV-3121 D-1 SHELL
AF-3123 FLOW SHEET
DWG NO
REFERENCE

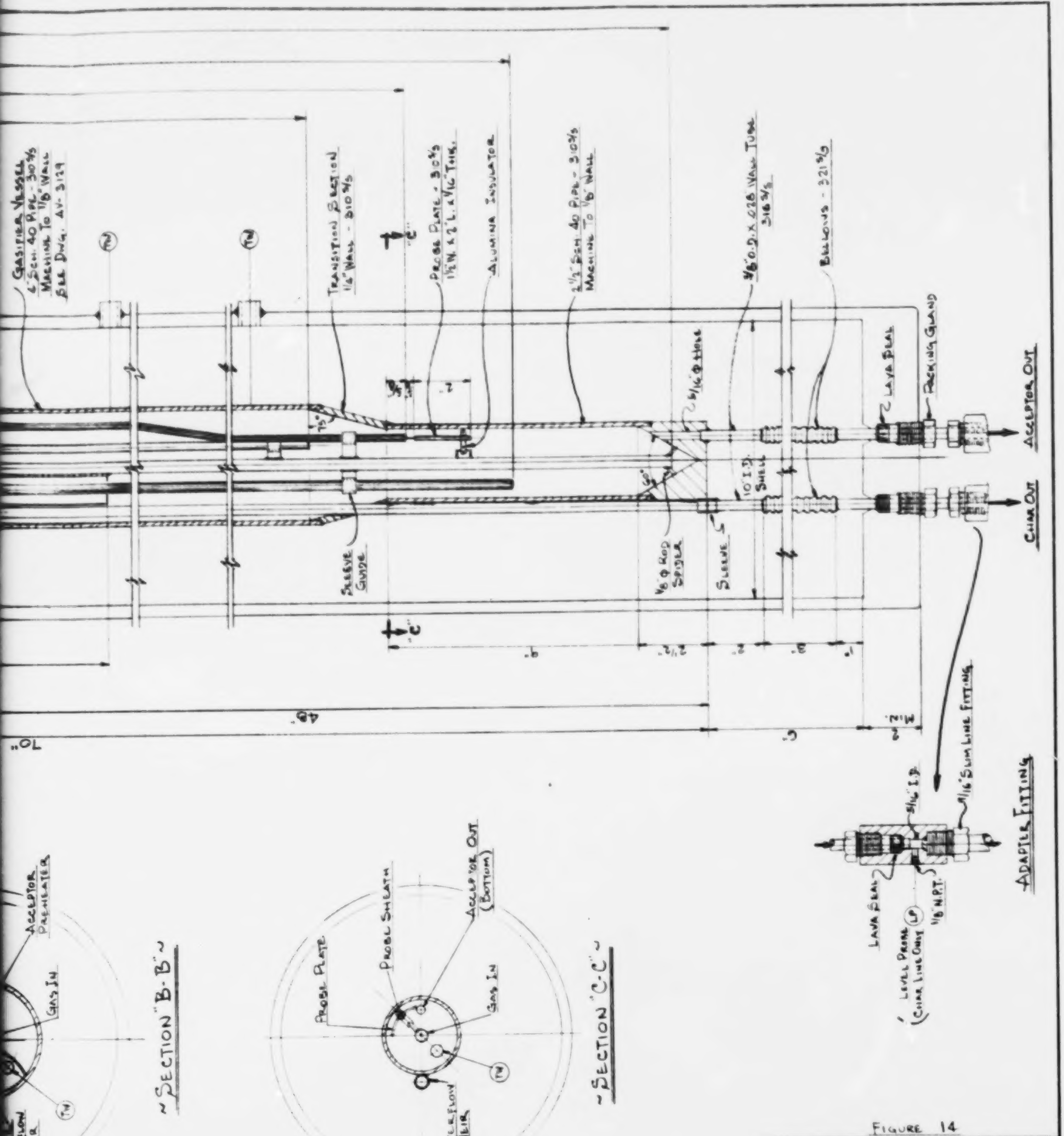

GASIFIER VESSEL 6" SCH. 40 PIPE - 304 S/S MACHINE TO 1/8" WALL SEE DWG. AV-3129
TW
TRANSITION SECTION 1/16" WALL - 310 S/S
PROBE PLATE - 304 S/S 1 1/8"W. x 2"L. x 1/16" THK.
ALUMINA INSULATOR
2 1/2" SCH. 40 PIPE - 310 S/S MACHINE TO 1/8" WALL
3/8" O.D. x .028 WALL TUBE 316 S/S
BELLOWS - 321 S/S
LAVA SEAL
PACKING GLAND
ACCEPTOR OUT
CHAR OUT
5/16 Φ HOLE
75°
60°
1/8 Φ ROD SPIDER
SLEEVE
SLEEVE GUIDE
10" I.D. SHELL
C
9"
2 1/2"
2"
3"
1"
48"
70"
6"
2" MIN.
ADAPTER FITTING
9/16" SLIM LINE FITTING
5/16" I.D.
1/8" N.P.T.
LEVEL PROBE (CHAR LINE ONLY)
LP
LAVA SEAL
ACCEPTOR PREHEATER
GAS IN
OVERFLOW WEIR
~SECTION "B-B"~
PROBE PLATE
PROBE SHEATH
ACCEPTOR OUT (BOTTOM)
GAS IN
TW
OVERFLOW WEIR
~SECTION "C-C"~

FIGURE 14

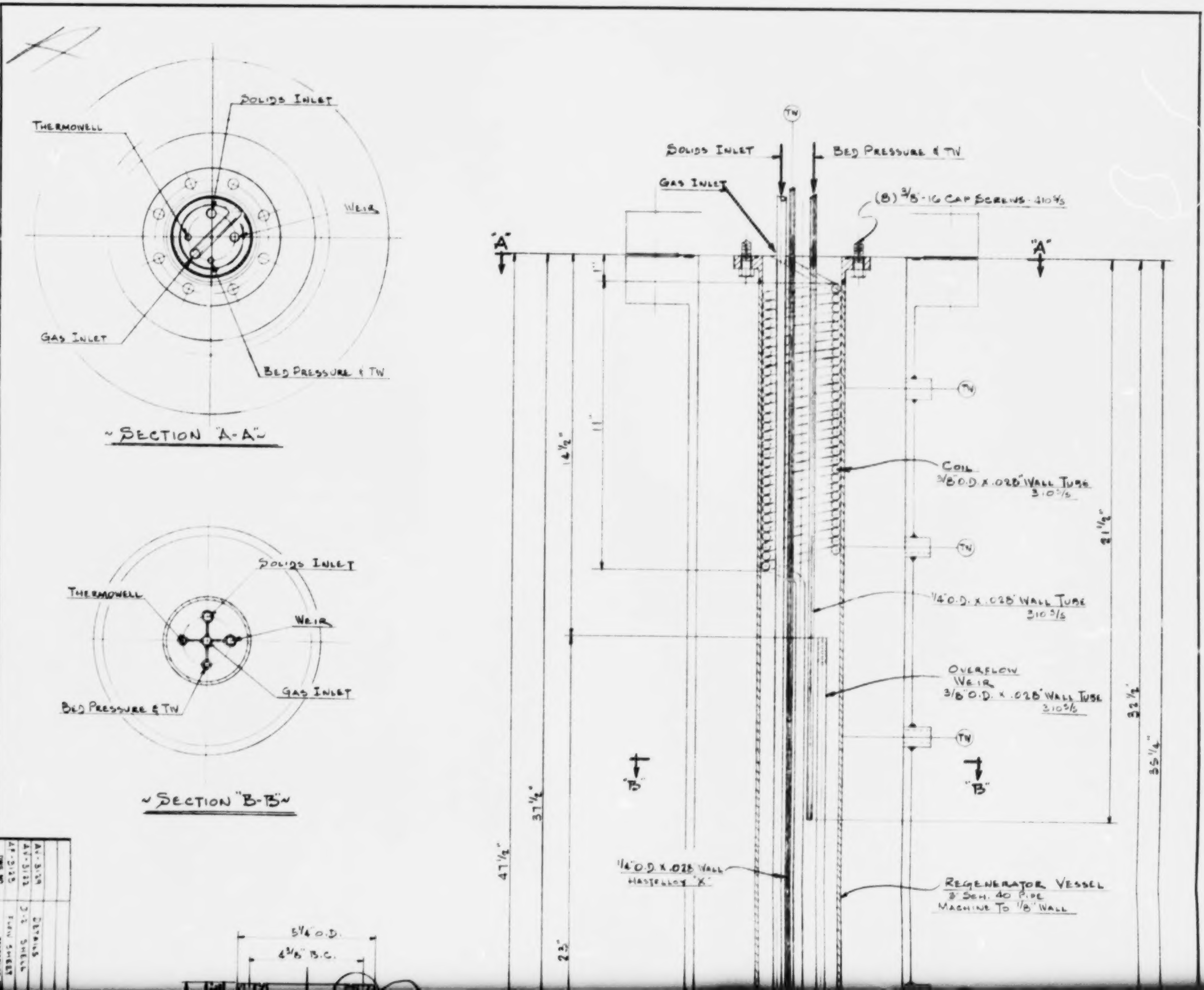
CONSOLIDATION COAL COMPANY
RESEARCH DIVISION
LIBRARY, PENNA

D-1 GASIFIER ASSEMBLY
CONTINUOUS CO_2 ACCEPTOR UNIT

AV-3125	DETAILS						
AV-3121	D-1 SHELL						
AF-3123	FLOW SHEET	1	8-6-71	ADDED BELLOWS TO TAIL PIECES			
DWG NO	REFERENCE DRAWINGS	NO	DATE	REVISION	BY	CHK	APP

DATE 11-25-70
DRN W.A.S.
CHK
APP
ISSUED FOR CONST

SCALE	PROJECT NO	DRAWING NUMBER	REV
1/2" = 1"	551.00	AV-3125	1


Solids Inlet
Thermowell
Weir
Gas Inlet
Bed Pressure & TW
~Section "A-A"~
Solids Inlet
Thermowell
Weir
Bed Pressure & TW
Gas Inlet
~Section "B-B"~
5¼" O.D.
4⅜" B.C.
TW
Solids Inlet
Bed Pressure & TW
Gas Inlet
(8) ⅜-16 Cap Screws · 310 S/S
"A"
"A"
1"
11"
14½"
37½"
47½"
23"
Coil
⅜" O.D. x .028" Wall Tube
310 S/S
TW
TW
TW
¼" O.D. x .028" Wall Tube
310 S/S
Overflow
Weir
⅜" O.D. x .028" Wall Tube
310 S/S
21½"
32½"
35¼"
"B"
"B"
¼" O.D. x .028" Wall
Hastelloy "X"
Regenerator Vessel
3" Sch. 40 Pipe
Machine To ⅛" Wall

DWG NO	REFERENCE
AV-3129	Details
AV-3122	J-2 Shell
AF-3123	Flow Sheet

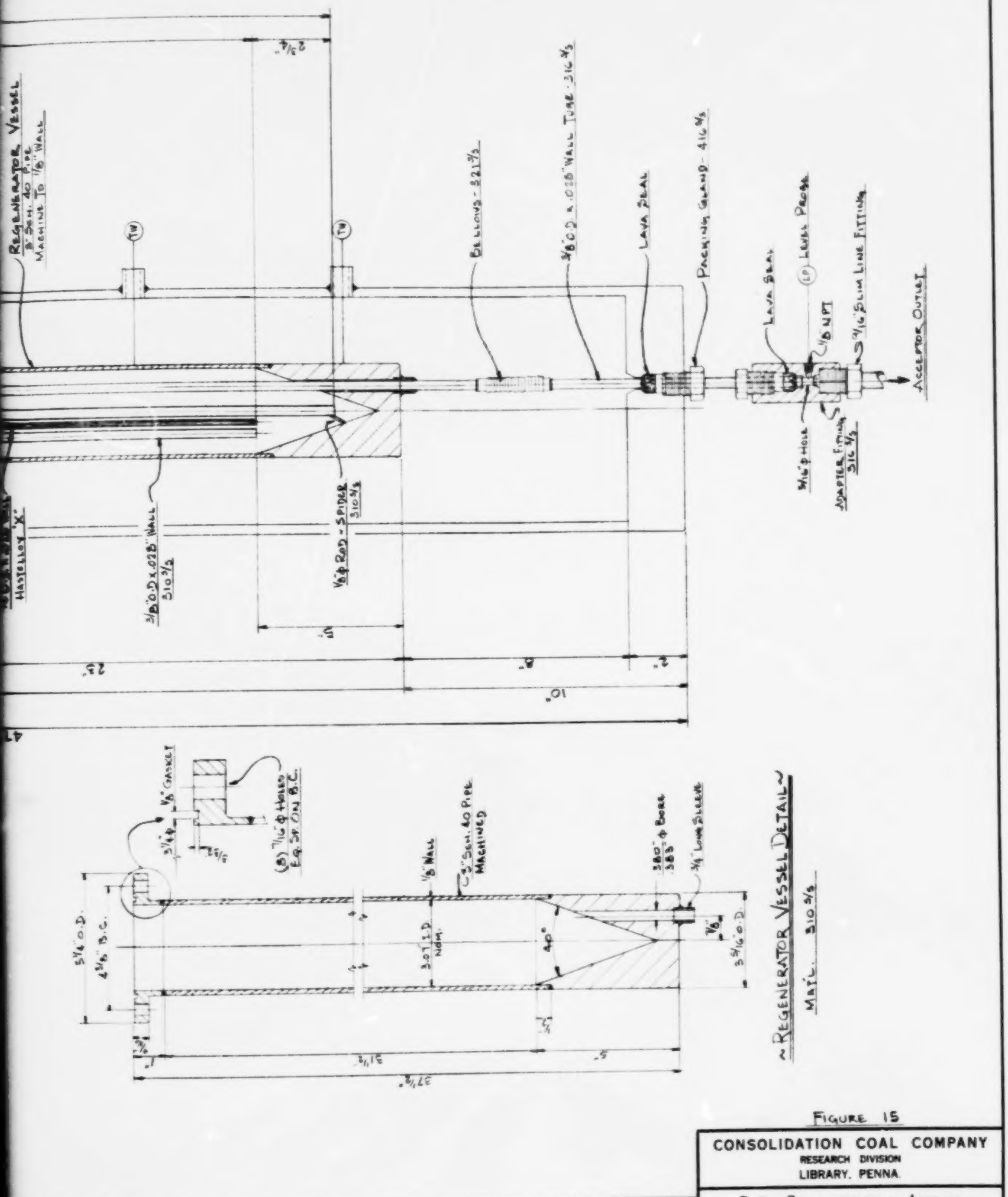

REGENERATOR VESSEL
8" SCH. 40 PIPE
MACHINE TO 1/8" WALL
TW
TW
BELLOWS - 321 S/S
3/8" O.D. x .028" WALL TUBE - 316 S/S
LAVA SEAL
PACKING GLAND - 416 S/S
LAVA SEAL
LP
LEVEL PROBE
1/8" NPT
9/16" SLIM LINE FITTING
ACCEPTOR OUTLET
5/16" φ HOLE
ADAPTER FITTING
316 S/S
1/8" φ ROD - SPIDER
310 S/S
3/8" O.D. x .028" WALL
310 S/S
HASTELLOY 'X'
2 1/2"
5"
25"
8"
2"
10"
REGENERATOR VESSEL DETAIL
MAT'L. 310 S/S
1/8" GASKET
(8) 7/16" φ HOLES
EQ. SP. ON B.C.
3/16" R.
3" SCH. 40 PIPE
MACHINED
1/8" WALL
3.07" I.D.
NOM.
40°
.380"
.385" φ BORE
3/4" LONG SLEEVE
1/8"
3 9/16" O.D.
5 1/4" O.D.
4 3/8" B.C.
1"
31 1/2"
5"
37 1/2"

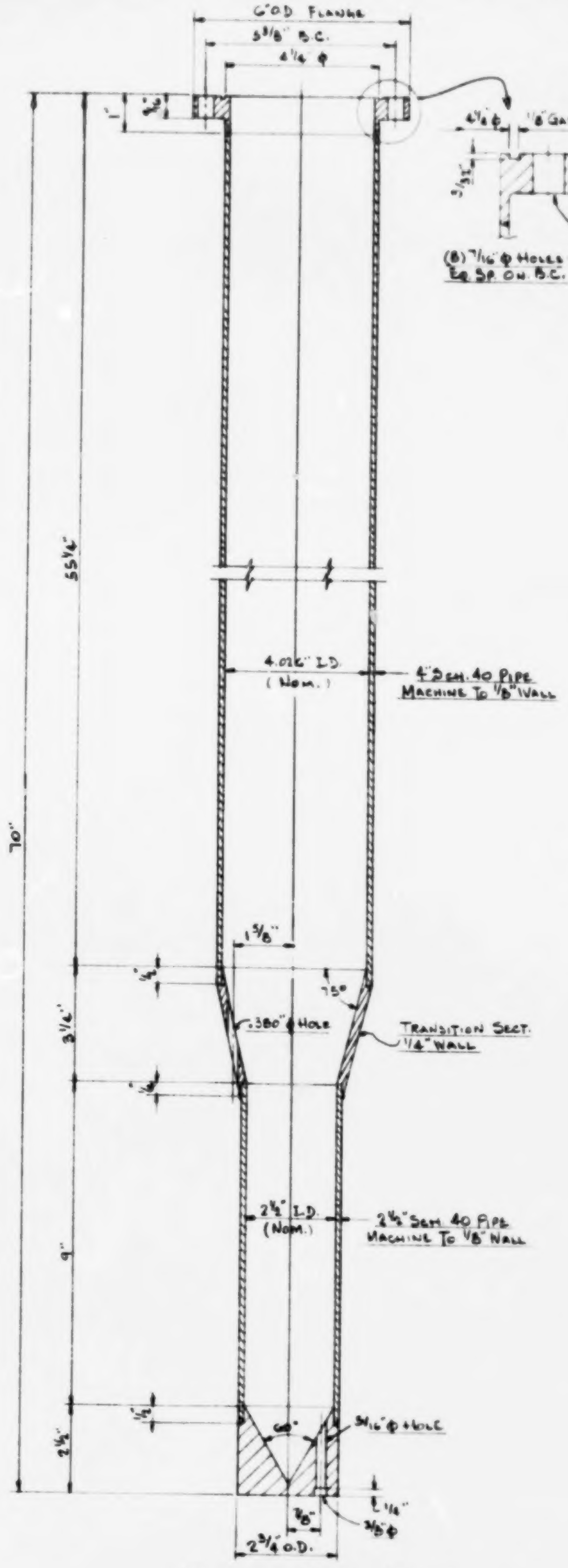
FIGURE 15

CONSOLIDATION COAL COMPANY
RESEARCH DIVISION
LIBRARY, PENNA.

D-2 REGENERATOR ASSEMBLY
CONTINUOUS CO_2 ACCEPTOR UNIT

								DATE	11-20-70
								DRN	W.A.S.
3129	DETAILS							CHK	
3122	D-2 SHELL							APP	
135	FLOW SHEET							ISSUED FOR CONST	
NO	REFERENCE DRAWINGS	NO	DATE	REVISION	BY	CHK	APP		

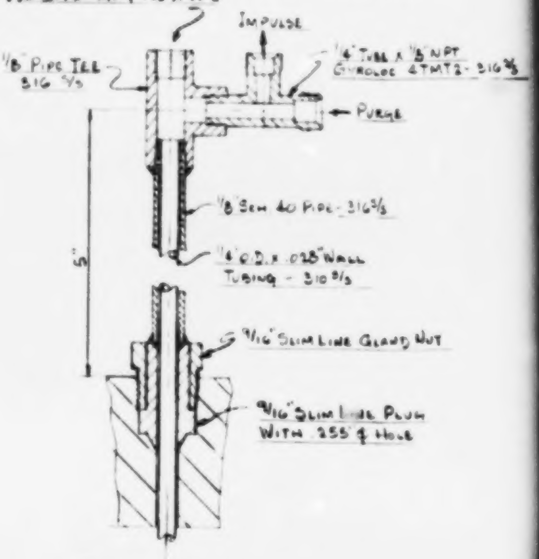
SCALE	PROJECT NO	DRAWING NUMBER	REV
1/2" = 1"	551.00	AV-3124	


6" O.D. FLANGE
5 3/8" B.C.
4 1/4" ⌀
4 1/4" ⌀
1/8" GASKET
3/32"
(8) 7/16" ⌀ HOLES
EQ. SP. ON B.C.
1"
3/4"
55 1/4"
70"
4.026" I.D.
(NOM.)
4" SCH. 40 PIPE
MACHINE TO 1/8" WALL
1 5/8"
3/4"
75°
.380" ⌀ HOLE
TRANSITION SECT.
1/4" WALL
3 1/4"
3/4"
2 1/2" I.D.
(NOM.)
2 1/2" SCH 40 PIPE
MACHINE TO 1/8" WALL
9"
1/2"
2 1/2"
60°
5/16" ⌀ HOLE
1/16"
3/8"
3/8" ⌀
2 3/4" O.D.

GASIFIER VESSEL DETAIL

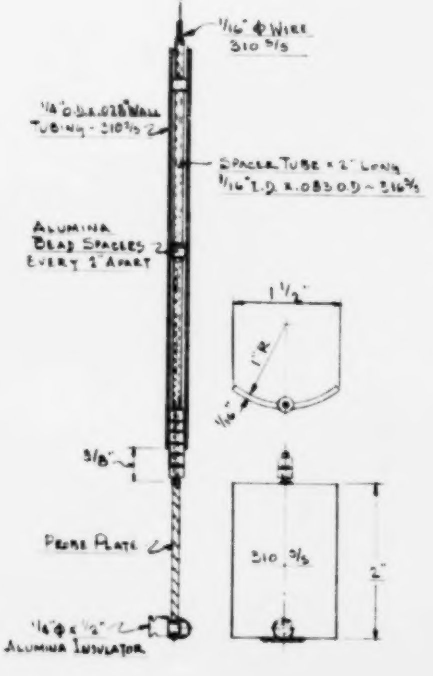
MATERIAL: ALL 310 S/S

PROBE - USE CONAX FITTING - EG-093-A-SS-L (WITHOUT CONDUCTOR)
ΔP TAP & TW - USE CONAX - MPG-125-A-SS-L


IMPULSE
1/8" PIPE TEE
316 S/S
1/4" TUBE x 1/8" NPT
GYROLOK 4TMT2 - 316 S/S
PURGE
1/8" SCH 40 PIPE - 316 S/S
1/4" O.D. x .028" WALL
TUBING - 310 S/S
5"
9/16" SLIM LINE GLAND NUT
9/16" SLIM LINE PLUG
WITH .255 ⌀ HOLE

TYPICAL DETAIL
GASIFIER PROBE SHEATH
REGENERATOR BED PRESSURE TAP & TW

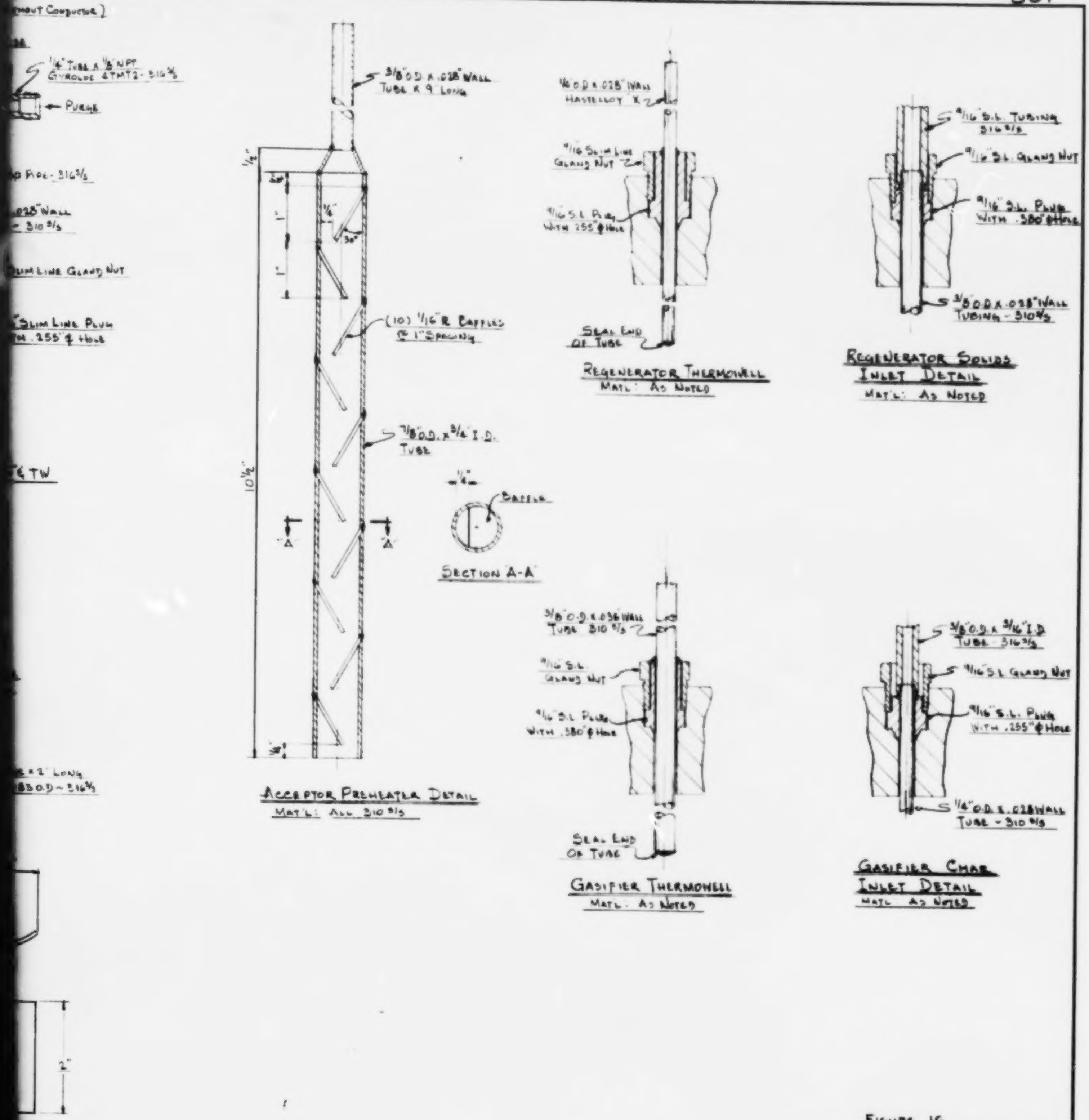
MATERIAL: AS NOTED


1/16" ⌀ WIRE
310 S/S
1/4" O.D. x .028" WALL
TUBING - 310 S/S
SPACER TUBE x 2" LONG
1/16" I.D. x .083 O.D. ~ 316 S/S
ALUMINA
BEAD SPACERS
EVERY 2" APART
1 1/2"
1" R
1/16"
3/8"
PROBE PLATE
310 S/S
2"
1/4" ⌀ x 1/2"
ALUMINA INSULATOR

GASIFIER PROBE DETAIL

MATERIAL: AS NOTED

AV-3129	DET
AV-3128	GASI
AV-3124	REG
AV-3122	D-2 REG
AV-3121	D-1 GAS
DWG NO	



THOUT CONDUCTOR)
1/4" TUBE x 1/8" NPT GYROLOC 4TMT2 - 316 S/S
PURGE
PIPE - 316 S/S
.028" WALL - 310 S/S
SLIM LINE GLAND NUT
" SLIM LINE PLUG WITH .255" ϕ HOLE
& TW
x 2" LONG .083 O.D. ~ 316 S/S
2"
3/8" O.D. x .028" WALL TUBE x 9" LONG
1/2"
1"
1"
1/8"
30°
(10) 1/16"R BAFFLES @ 1" SPACING
7/8" O.D. x 3/4" I.D. TUBE
10 1/2"
"A"
"A"
1/8"
BAFFLE
SECTION "A-A"
ACCEPTOR PREHEATER DETAIL
MAT'L: ALL 310 S/S
1/8" O.D. x .028" WALL HASTELLOY X
9/16 SLIM LINE GLAND NUT
9/16 S.L. PLUG WITH .255" ϕ HOLE
SEAL END OF TUBE
REGENERATOR THERMOWELL
MAT'L: AS NOTED
9/16" S.L. TUBING 316 S/S
9/16" S.L. GLAND NUT
9/16" S.L. PLUG WITH .380" ϕ HOLE
3/8" O.D. x .028" WALL TUBING - 310 S/S
REGENERATOR SOLIDS INLET DETAIL
MAT'L: AS NOTED
3/8" O.D. x .036 WALL TUBE 310 S/S
9/16" S.L. GLAND NUT
9/16" S.L. PLUG WITH .380" ϕ HOLE
SEAL END OF TUBE
GASIFIER THERMOWELL
MAT'L. AS NOTED
3/8" O.D. x 3/16" I.D. TUBE - 316 S/S
9/16" S.L. GLAND NUT
9/16" S.L. PLUG WITH .255" ϕ HOLE
1/4" O.D. x .028 WALL TUBE - 310 S/S
GASIFIER CHAR INLET DETAIL
MAT'L AS NOTED

FIGURE 16

CONSOLIDATION COAL COMPANY
RESEARCH DIVISION
LIBRARY, PENNA.

DWG NO	REFERENCE DRAWINGS
AV-3129	DETAILS
AV-3125	GASIFIER ASSEMBLY
AV-3124	REGENERATOR ASSEMBLY
AV-3122	D-2 REGENERATOR SHELL
AV-3121	D-1 GASIFIER SHELL

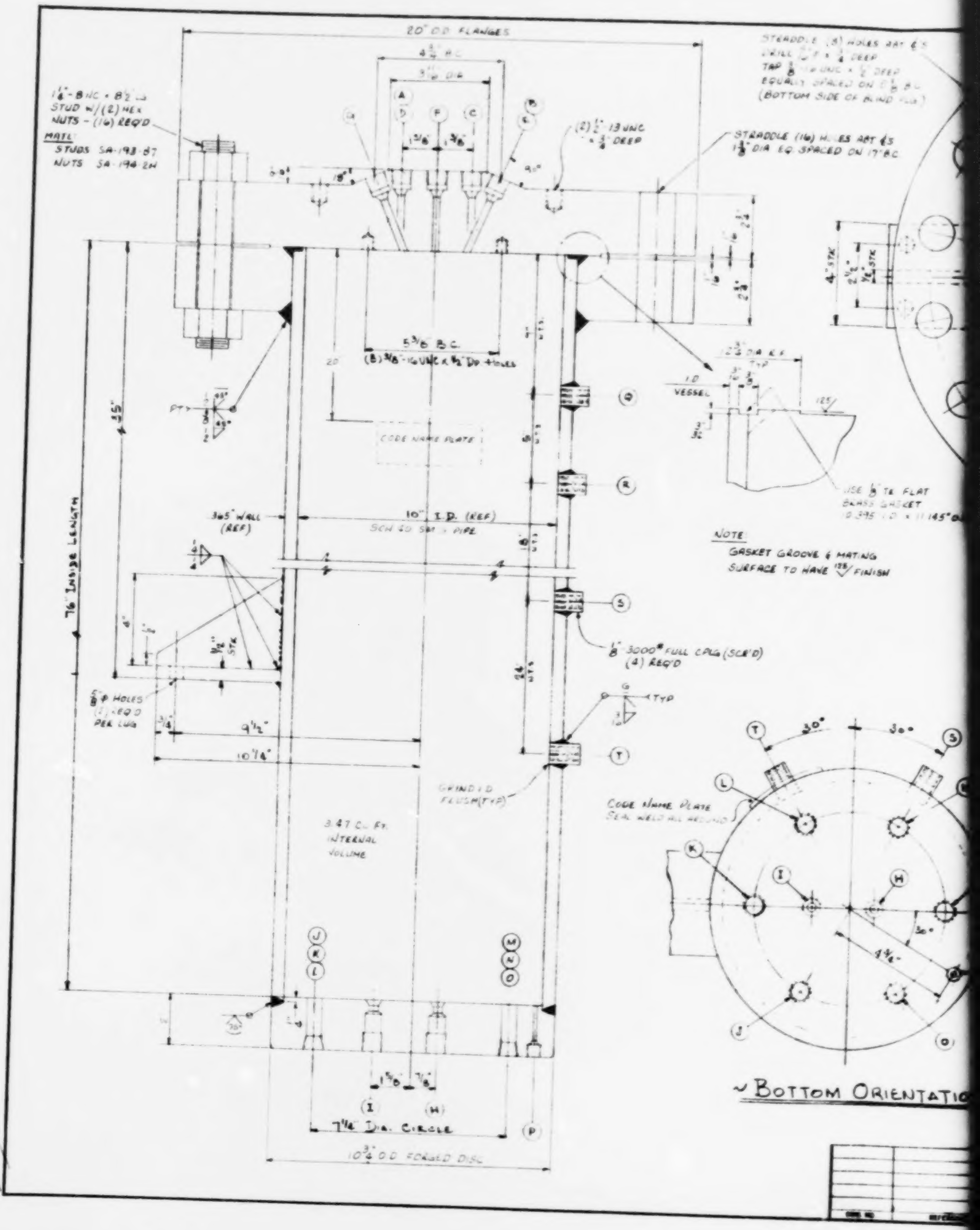
NO	DATE	REVISION	BY	CHK	APP

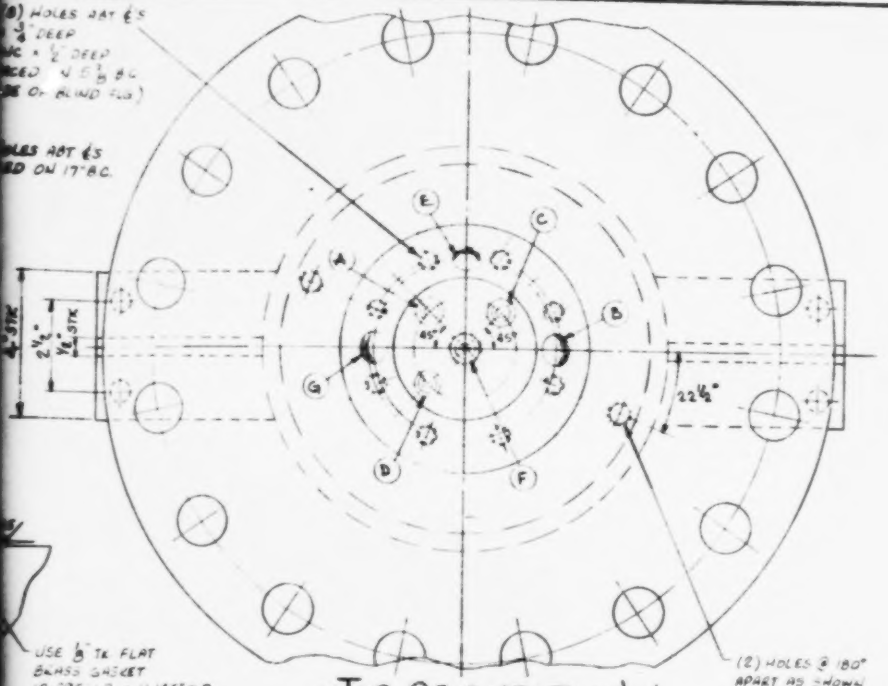
DATE WAS
DRN 12-18-70
CHK
APP
ISSUED FOR CONST

DETAILS FOR GASIFIER & REGENERATOR
FITTINGS & INTERNALS
CONTINUOUS CO_2 ACCEPTOR UNIT

SCALE	PROJECT NO	DRAWING NUMBER	REV
1/2 & FULL	881.00	AV-3129	

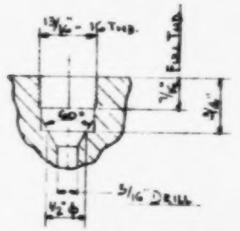
B


20" O.D. FLANGES
4 3/4" B.C.
3 11/16" DIA
A
B
C
D
E
F
G
1 3/16"
1 3/8"
1 1/8" - 8 NC × 8 1/2" LG
STUD W/(2) HEX
NUTS - (16) REQ'D
MAT'L
STUDS SA-193-B7
NUTS SA-194-2H
(2) 1/2"-13 UNC × 3/4" DEEP
90°
STRADDLE (8) HOLES ABT ℄S
DRILL 5/16" Ø × 3/4" DEEP
TAP 3/8"-16 UNC × 1/2" DEEP
EQUALLY SPACED ON 5 3/8" B.C.
(BOTTOM SIDE OF BLIND FLG)
STRADDLE (16) HOLES ABT ℄S
1 3/8" DIA EQ. SPACED ON 17" B.C.
5 3/8" B.C.
(8) 3/8"-16 UNC × 1/2" DP. HOLES
20
CODE NAME PLATE
76" INSIDE LENGTH
.365" WALL (REF)
10" I.D. (REF)
SCH 60 SMLS PIPE
4" STK
12 3/4" DIA R.F. TYP
I.D. VESSEL
3/16"
3/8"
3/32"
125
USE 1/8" TK FLAT BRASS GASKET 10.395 I.D. × 11.145" O.D.
NOTE
GASKET GROOVE & MATING SURFACE TO HAVE 125 FINISH
Q
R
S
T
1/8" 3000# FULL CPLG (SCR'D) (4) REQ'D
G TYP
5/8" Ø HOLES (2) REQ'D PER LUG
9 1/2"
10 1/4"
3/4"
GRIND I.D. FLUSH (TYP)
3.47 Cu. Ft. INTERNAL VOLUME
J
K
L
M
N
O
P
H
I
1 5/8"
7/8"
7 1/4" DIA. CIRCLE
10 3/4" O.D. FORGED DISC
30°
CODE NAME PLATE SEAL WELD ALL AROUND
4 1/8"
BOTTOM ORIENTATIO


(8) HOLES ABT ℄S
3/8" DEEP
NC × 1/2" DEEP
CED ON 5 1/8 BC
ES OF BLIND FLG)
LES ABT ℄S
ED ON 17" B.C.
4" STK
2 1/2"
1/2" STK
A
B
C
D
E
F
G
45°
45°
22 1/2°
USE 1/8" TK FLAT
BRASS GASKET
10.395" I.D. × 11.145" O.D.
4 MATING
125 ✓ FINISH
(2) HOLES @ 180°
APART AS SHOWN
ON 9" B.C.
DRILL 7/16" ⌀ × 1" DEEP
TAP 1/2"-13 UNC × 3/4" DEEP
(FOR LIFTING LUGS)

~TOP ORIENTATION~

SCHEDULE OF OPENINGS		
ITEM	DESCRIPTION	SERVICE
A	9/16" SLIM LINE	CHAR INLET
B	9/16" SLIM LINE	GAS OUTLET
C	9/16" SLIM LINE	THERMOWELL
D	9/16" SLIM LINE	PROBE
E	1/4" H.P. CONE	PRESSURE TAP
F	9/16" SLIM LINE	GAS INLET
G	9/16" SLIM LINE	ACCEPTOR INLET
H	3/4" SPECIAL	CHAR OUTLET
I	3/4" SPECIAL	ACCEPTOR OUTLET
J	1/2" PIPE TAP	CONAX ELECT CONNECTIONS
K	1/2" PIPE TAP	
L	1/2" PIPE TAP	
M	1/2" PIPE TAP	
N	1/2" PIPE TAP	
O	1/2" PIPE TAP	
P	1/4" H.P. CONE	BALANCE GAS INLET
Q	1/8" PIPE COUPLING	THERMOWELL
R	1/8" PIPE COUPLING	THERMOWELL
S	1/8" PIPE COUPLING	THERMOWELL
T	1/8" PIPE COUPLING	THERMOWELL


13/16" - 16 THD.
FULL THD.
7/16"
3/4"
60°
5/16" DRILL
1/2" ⌀

TYPICAL DETAIL FOR Ⓐ Ⓑ Ⓒ Ⓓ Ⓕ Ⓖ

SPECIFICATIONS:

DESIGN PRESSURE 450 PSIG
DESIGN TEMPERATURE 900°F. (INCL. CLOSURES)
MATERIAL 316 S/S UNLESS NOTED
SHELL TO BE FABRICATED FROM SEAMLESS PIPE

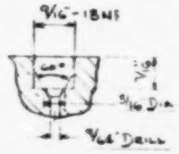
NOTES:

1. MANUFACTURE & STAMP IN ACCORDANCE WITH ASME PRESSURE VESSEL CODE SECT. VIII & PENNA CODE
2. TEST PRESSURE

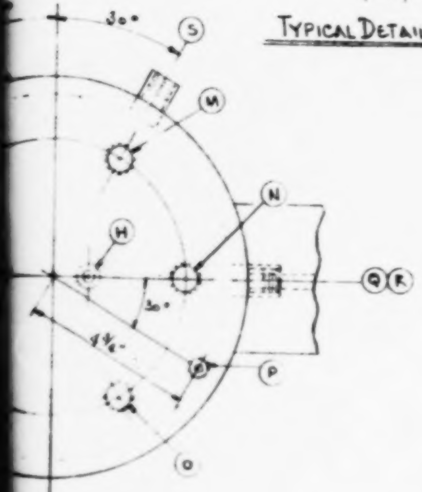
$$450 \times 1.5 \times \frac{18,750}{15,550} = 817 \text{ PSIG}$$

U
NATIONAL BOARD Nº 286
MANUFACTURED BY IONICS
DIV 1 MAX ALLOW WORKING PRESS. 450 PSI @ 900°F
MFR'S SERIAL Nº 4786-1
YEAR BUILT 1971
SHELL TK 0.365 NOM.
BOT HD TK 2"

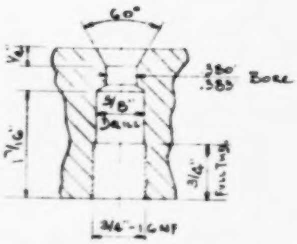
CODE NAME PLATE


9/16" - 18 NF
1/16"
60°
5/16 DIA.
9/64" DRILL

TYPICAL DETAIL FOR Ⓔ Ⓟ


30°
S
M
N
H
Q R
30°
9 1/4"
P
O

~BOTTOM ORIENTATION~


60°
1/4"
.380
.383 BORE
3/8" DRILL
1 7/16"
3/4" FULL THD.
3/4"-16 NF

TYPICAL DETAIL FOR Ⓗ Ⓘ

(1) UNIT REQ'D
PROJ. 551.00

3 NOTE 1 REVISED	PFD 2-2-71
2 TEST PRESSURE WAS 542 PSIG	PFD 1-27-71
1 PER CUST COMMENTS	1/20/71 TA
REVISIONS	

NO.	DATE	REVISION	BY	CHK	APP
~~1~~	~~11/24/70~~	~~CHANGED DIM. OF Ⓠ Ⓡ TO MATCH SIMILAR VESSEL~~	~~WAS~~		

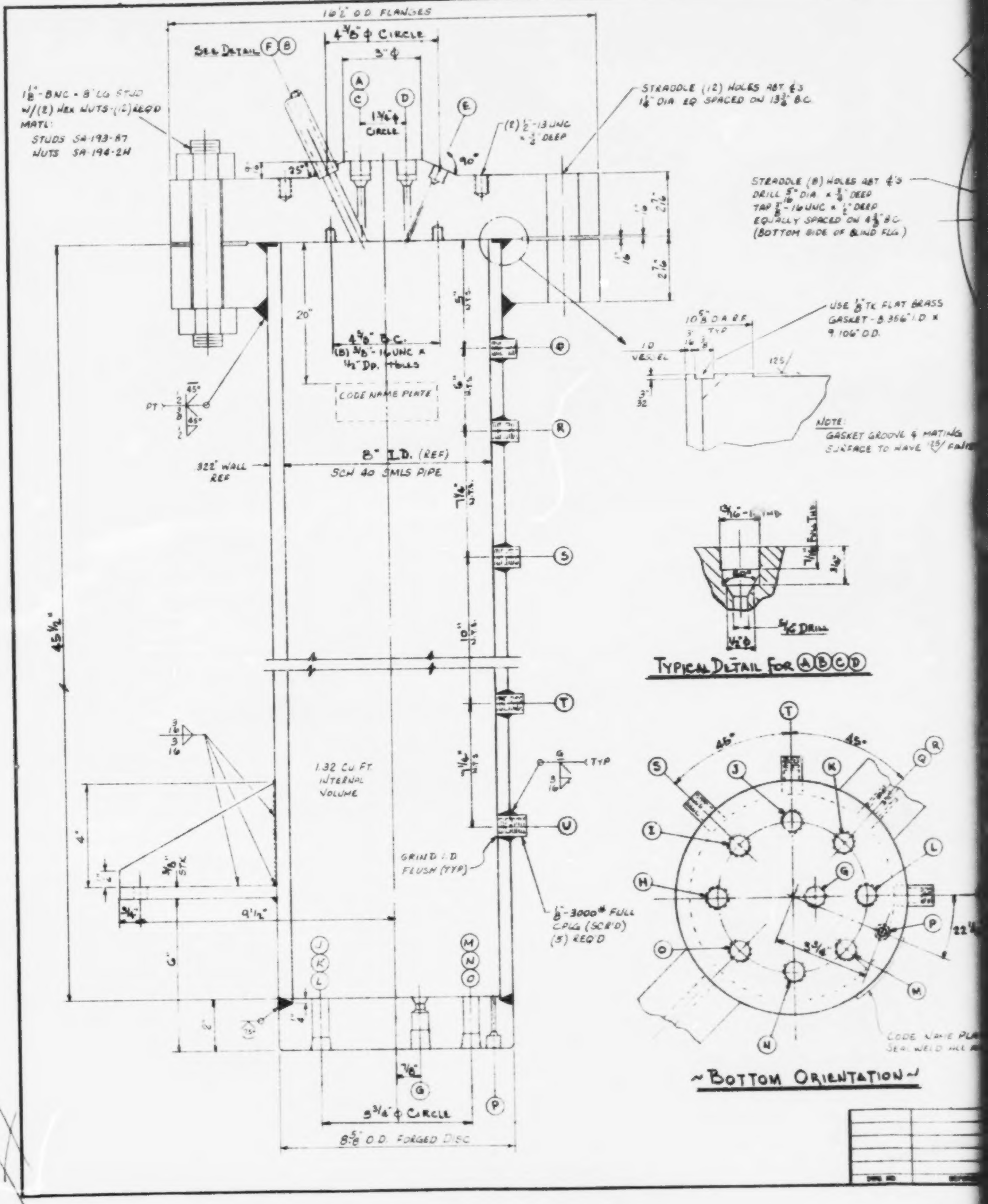
DWG NO | REFERENCE DRAWINGS

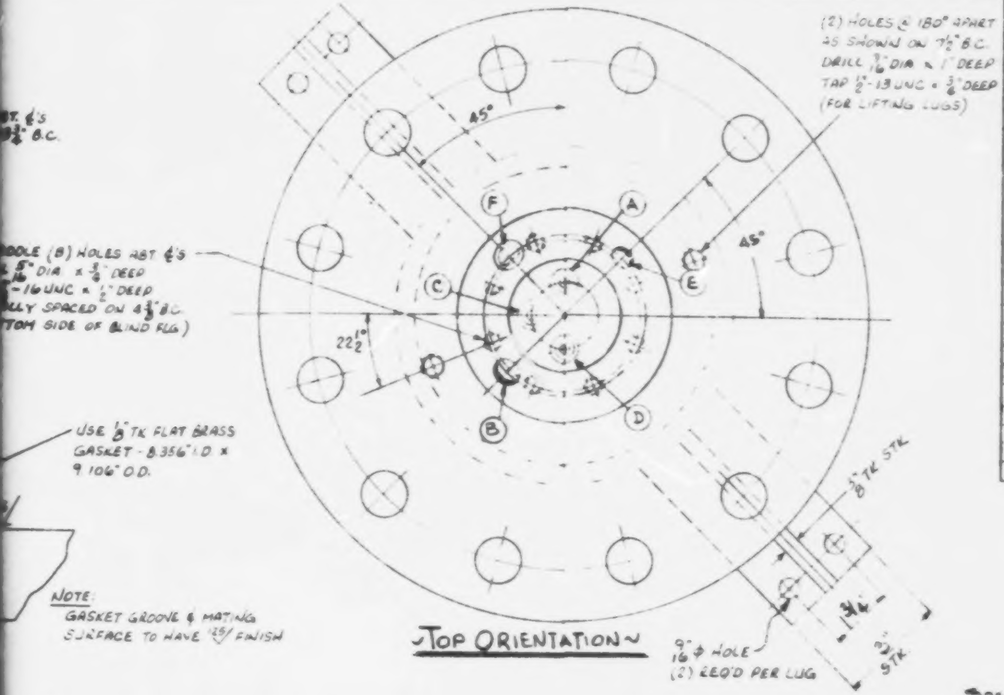
DATE 11-12-70
DRN W.A.S.
CHK
APP
ISSUED FOR CONST.

IONICS. INCORPORATED
BRIDGEVILLE PA

FOR CONSOLIDATION COAL CO. INC
P.O. 09001-551 | REF DWG AV-3121
D-1 GASIFIER SHELL
CONTINUOUS CO2 ACCEPTOR UNIT
SCALE | DR PFD 1-12-71
| AP DMF 1-13-71
DWG NO. D-4786-1 | REV 3

FIGURE 11

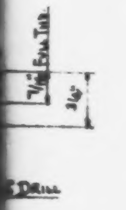

16½" O.D. FLANGES
4⅞" ⌀ CIRCLE
3" ⌀
SEE DETAIL Ⓕ Ⓑ
Ⓐ
Ⓒ
Ⓓ
1⁹⁄₁₆" ⌀ CIRCLE
Ⓔ
1⅛"-8NC × 8" LG STUD W/(2) HEX NUTS-(12) REQ'D
MAT'L:
STUDS SA-193-B7
NUTS SA-194-2H
25°
90°
(2) ½"-13 UNC × ¾" DEEP
STRADDLE (12) HOLES ABT ℄'S 1¼" DIA EQ SPACED ON 13¾" B.C.
STRADDLE (8) HOLES ABT ℄'S DRILL ⁵⁄₁₆" DIA × ⅝" DEEP TAP ⅜"-16 UNC × ½" DEEP EQUALLY SPACED ON 4⅞" B.C. (BOTTOM SIDE OF BLIND FLG.)
2 ⁷⁄₁₆"
1⁄16"
2 ⁷⁄₁₆"
20"
4⅞" B.C.
(8) ⅜"-16 UNC × ½" DP. HOLES
CODE NAME PLATE
5" 2 T.S.
6" 2 T.S.
Ⓠ
Ⓡ
PT
45°
45°
.322" WALL REF
8" I.D. (REF)
SCH 40 SMLS PIPE
7⁹⁄₁₆" 2 T.S.
Ⓢ
10" 2 T.S.
Ⓣ
7⁹⁄₁₆" 2 T.S.
Ⓤ
45½"
1.32 CU FT INTERNAL VOLUME
TYP
GRIND I.D FLUSH (TYP)
⅛"-3000# FULL CPLG (SCR'D) (5) REQ'D
4"
⅜" STK
9½"
6"
2"
Ⓙ
Ⓚ
Ⓛ
Ⓜ
Ⓝ
Ⓞ
⅛"
Ⓖ
Ⓟ
5¾" ⌀ CIRCLE
8⅝" O.D. FORGED DISC
I.D VESSEL
10⅝" O.A. R.F. TYP
3⁄16
⅜
3⁄32
125
USE ⅛" TK FLAT BRASS GASKET - 8.356" I.D × 9.106" O.D.
NOTE: GASKET GROOVE & MATING SURFACE TO HAVE 125 FINISH
⁹⁄₁₆-18 UNF
7⁄16 FULL THD
1⁵⁄₁₆"
60°
⁵⁄₁₆ DRILL
½" ⌀
TYPICAL DETAIL FOR Ⓐ Ⓑ Ⓒ Ⓓ
45°
45°
Ⓣ
Ⓢ
Ⓙ
Ⓚ
Ⓠ Ⓡ
Ⓘ
Ⓛ
Ⓗ
Ⓖ
Ⓟ
Ⓞ
Ⓜ
Ⓝ
5¾"
22½
CODE NAME PLA
SEAL WELD ALL A
~BOTTOM ORIENTATION~
DWG NO.


(2) HOLES @ 180° APART
AS SHOWN ON 7½" B.C.
DRILL 9/16" DIA x 1" DEEP
TAP ½"-13 UNC x ⅜" DEEP
(FOR LIFTING LUGS)
45°
45°
A
B
C
D
E
F
22½°
..T. ¢'S
..⅜" B.C.
..DDLE (8) HOLES ABT ¢'S
.. 5/16" DIA x ¾" DEEP
..-16 UNC x ½" DEEP
..LLY SPACED ON 4⅜" B.C.
..TOM SIDE OF BLIND FLG.)
USE ⅛" TK FLAT BRASS
GASKET - 8.356" I.D. x
9.106" O.D.
NOTE:
GASKET GROOVE & MATING
SURFACE TO HAVE 125/ FINISH
⅝" TK STK
3/4"
3" STK
9/16" ⌀ HOLE
(2) REQ'D PER LUG

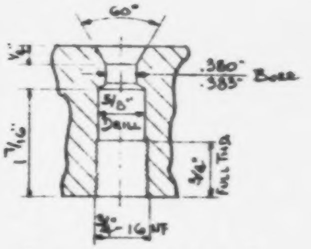
~TOP ORIENTATION~

SCHEDULE OF OPENINGS

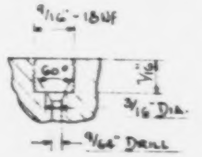
ITEM	DESCRIPTION	SERVICE
A	9/16" SLIM LINE	SOLIDS INLET
B	9/16" SLIM LINE	GAS INLET
C	9/16" SLIM LINE	THERMOWELL
D	9/16" SLIM LINE	BED PRESSURE TAP
E	1/4" H.P. CONE	PRESSURE TAP
F	1/2" SPECIAL	GAS OUTLET
G	3/4" SPECIAL	ACCEPTOR OUTLET
H	1/2" PIPE TAP	COVAX ELECT. CONNECTIONS
I	1/2" PIPE TAP	
J	1/2" PIPE TAP	
K	1/2" PIPE TAP	
L	1/2" PIPE TAP	
M	1/2" PIPE TAP	
N	1/2" PIPE TAP	
O	1/2" PIPE TAP	
P	1/4" H.P. CONE	BALANCE GAS INLET
Q	1/8" PIPE COUPLING	THERMOWELL
R	1/8" PIPE COUPLING	THERMOWELL
S	1/8" PIPE COUPLING	THERMOWELL
T	1/8" PIPE COUPLING	THERMOWELL
U	1/8" PIPE COUPLING	THERMOWELL


..FULL THD.
3/16"
..S DRILL

ⒶⒷⒸⒹ


60°
.380"
.385" BORE
5/16" DRILL
1/16"
3/4" FULL THD.
3/4"-16 UNF

TYPICAL DETAIL FOR Ⓖ


9/16"-18 UNF
60°
3/16" DIA.
9/64" DRILL

TYPICAL DETAIL FOR ⒺⓅ

SPECIFICATIONS:

DESIGN PRESSURE 450 PSIG

DESIGN TEMPERATURE 900°F (INCL. CLOSURE)

MATERIAL 316 S/S UNLESS NOTED

SHELL TO BE FABRICATED FROM SEAMLESS PIPE

NOTES:

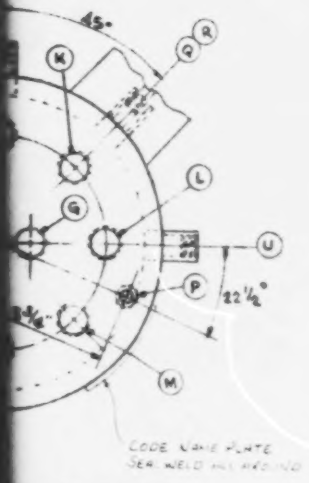
1. MANUFACTURE & STAMP IN ACCORDANCE WITH ASME PRESSURE VESSEL CODE SECT. VIII & PENNA CODE
2. TEST PRESSURE

$$450 \times 1.5 \times \frac{18.750}{15.550} = 817 \text{ PSIG}$$

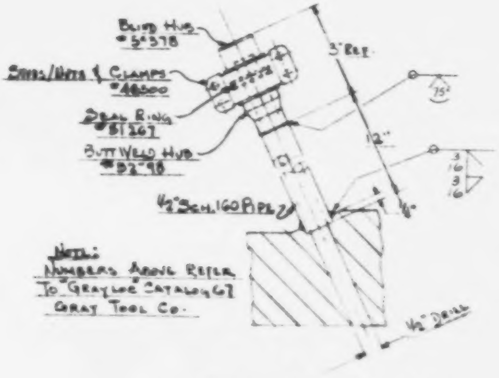
DIV. 1

NATIONAL BOARD NO 287
MANUFACTURED BY IONICS
MAX ALLOW WORKING PRESS 450 PSI @ 900°F
MFR'S SERIAL NO 4786-2
YEAR BUILT 1971
SHELL TK 0.322 NOM
BOT HD TK 2

CODE NAME PLATE


45°
K
Q
R
L
G
U
P
22½°
M
CODE NAME PLATE
SEAL WELD ALL AROUND

..RIENTATION~


BLIND HUB
#5"378
STUDS/NUTS & CLAMPS
#46500
SEAL RING
#51267
BUTT WELD HUB
#52"98
½" SCH. 160 PIPE
3" REF.
12"
75°
3/16
3/16
½" DRILL
NOTE:
NUMBERS ABOVE REFER
TO "GRAYLOC" CATALOG 67
GRAY TOOL CO.

TYPICAL DETAIL FOR Ⓕ

(1) UNIT REQ'D

PROJ. 551.00

5 NOTE 1 REVISED
4 ON CODE NAME & "U" SYMBOL WAS "UM" PED 2-12-71
3 ON CODE NAME & "UM" SYMBOL WAS "U" PED 1-29-71
2 TEST PRESSURE WAS 542 PSIG PED 1-27-71
1 PER CUST COMMENTS 1/26/71 TA
REVISIONS

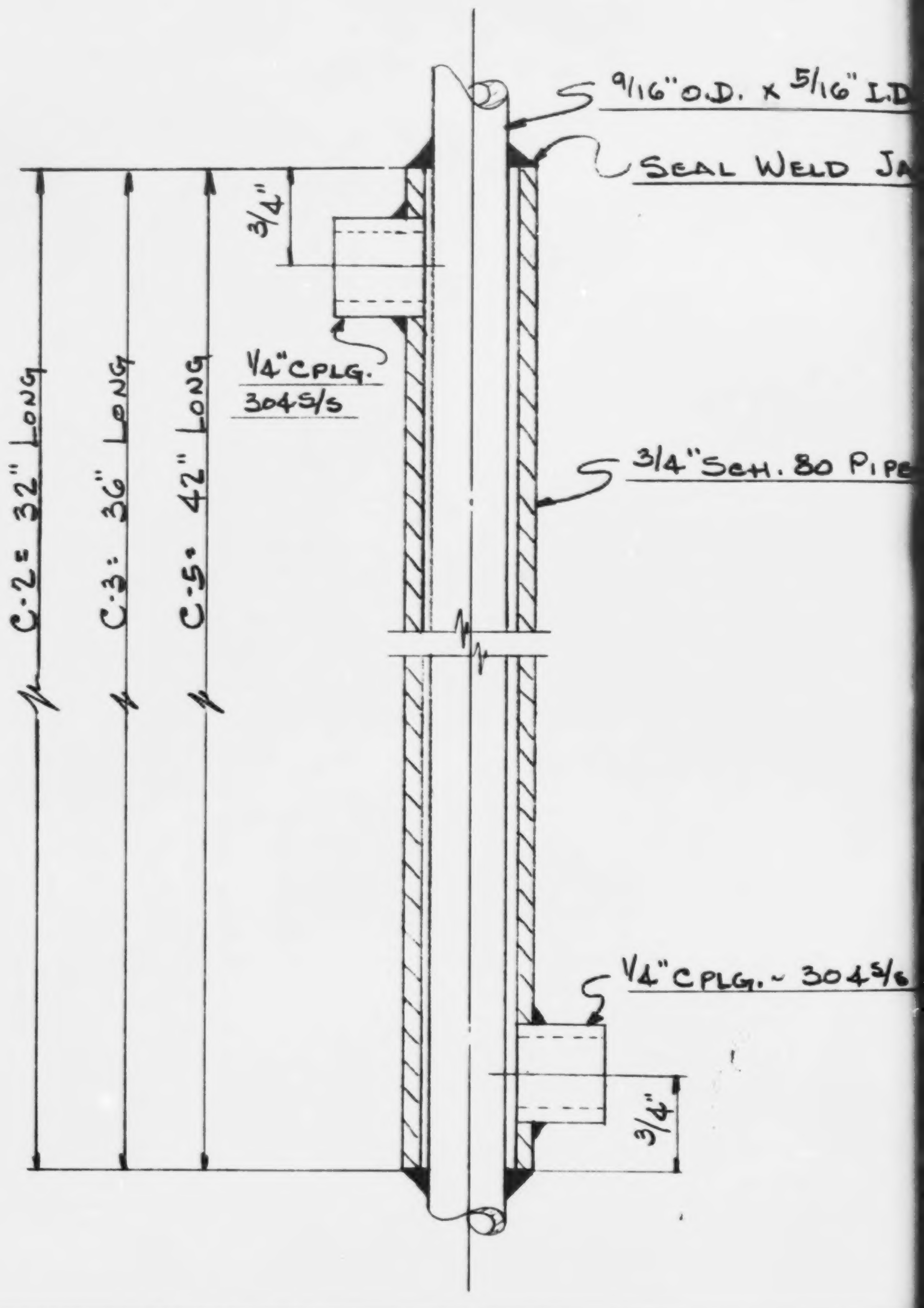
DWG. NO.	REFERENCE DRAWINGS	NO.	DATE	REVISION	BY	CHK.	APP.

DATE 11-12-70
DRN W.A.S.
CHK
APP
ISSUED FOR CONST

IONICS. INCORPORATED
BRIDGEVILLE, PA

FOR CONSOLIDATION COAL CO., INC
P.O. 090013-551 REF DWG AV-3122
D-2 REGENERATOR SHELL
CONTINUOUS CO2 ACCEPTOR UNIT
SCALE DR PED 1-12-71
CK APP [illegible]
DWG. NO. D-4786-2 REV. 4

FIGURE 18


9/16" O.D. x 5/16" I.D
SEAL WELD JA
3/4"
1/4" CPLG.
304 S/S
C-2 = 32" LONG
C-3 = 36" LONG
C-5 = 42" LONG
3/4" SCH. 80 PIPE
1/4" CPLG. ~ 304 S/S
3/4"

" I.D. TUBING (PROCESS LINE)

JACKET TO TUBING IN FIELD (TYP.)

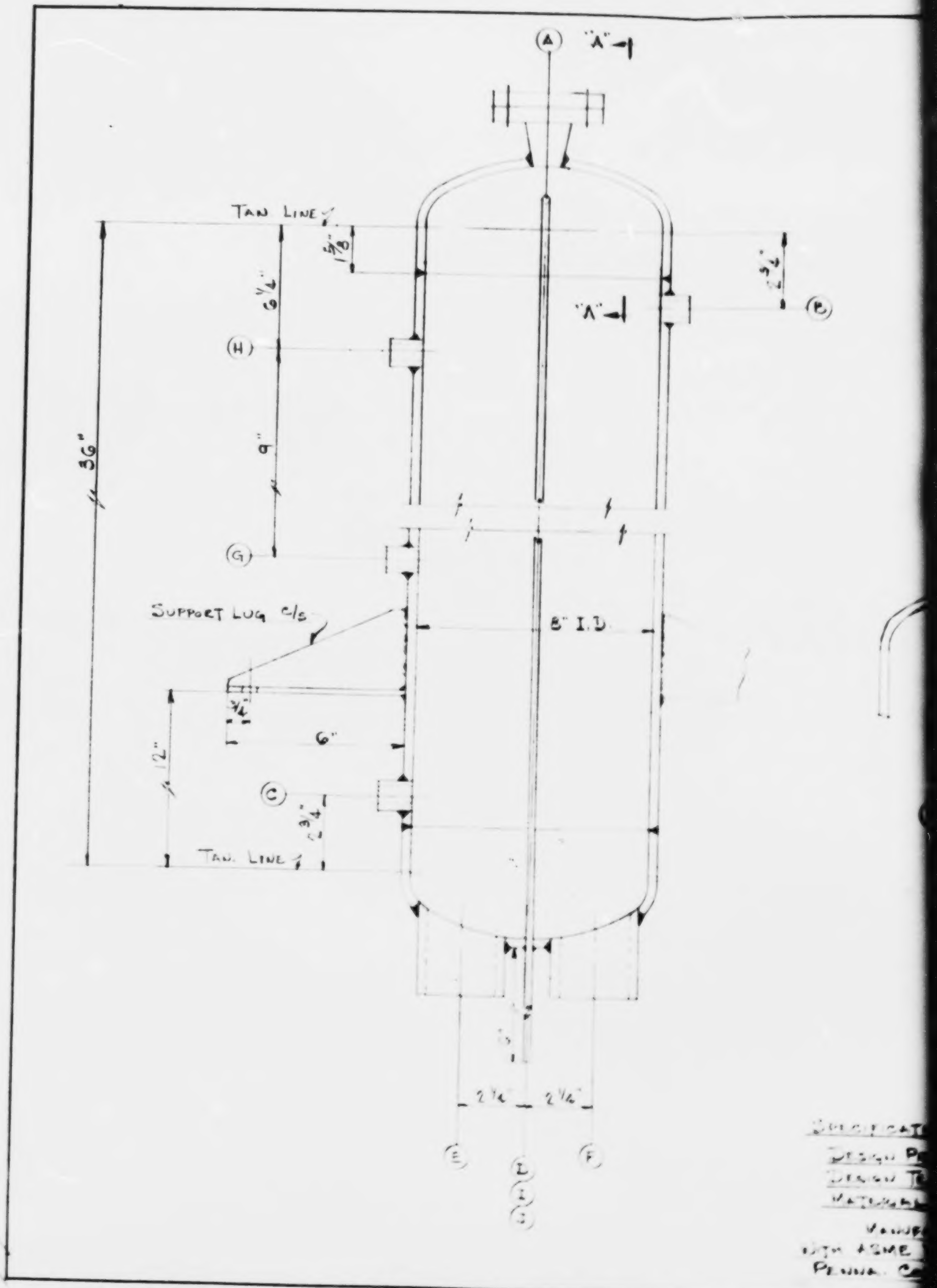
PIPE - 304 S/S

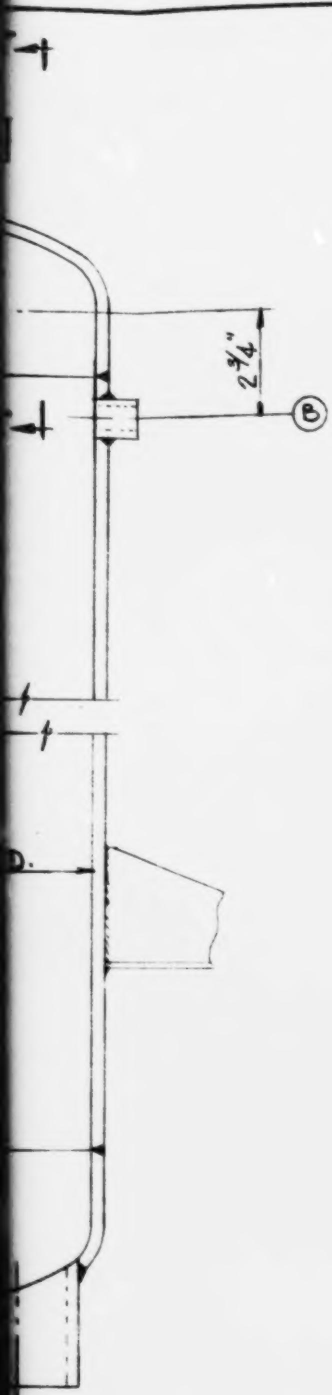
HYDROSTATIC TEST : 150 PSIG

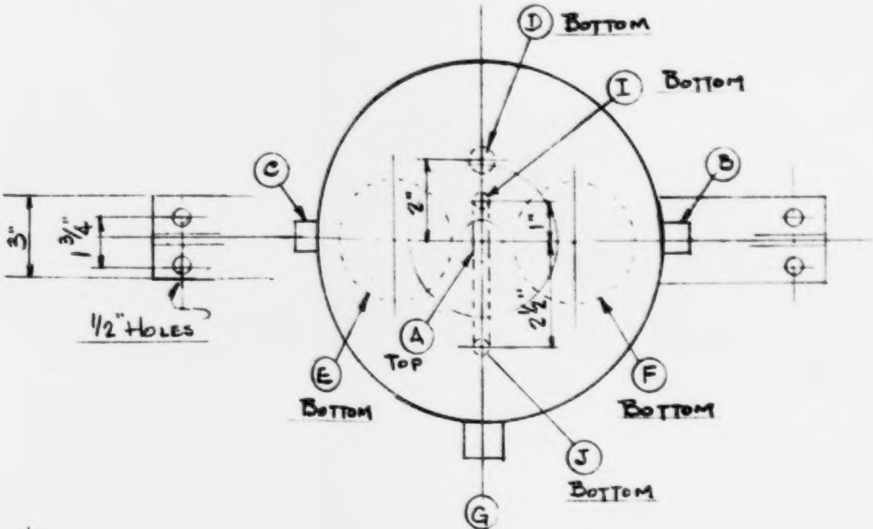
4 S/S

FIGURE 19

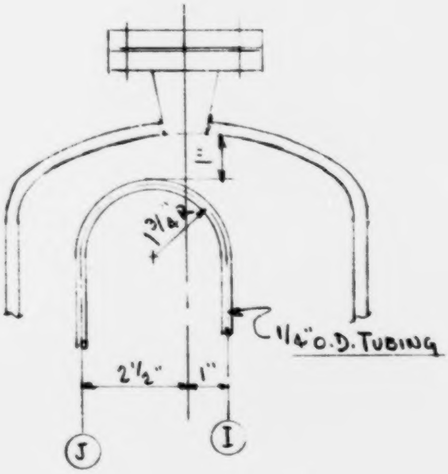
REVISIONS			CONSOLIDATION COAL CO. RESEARCH DIVISION LIBRARY, PENNA.		
			C-2, C-3 & C-5 COOLERS CONTINUOUS CO_2 ACCEPTOR GASIFICATION UNIT		
DRAWN	WS	6/29/65	SCALE	PROJECT	DRAWING NO.
CHKD.			FULL	531.03	CV-2709
APPR.					


A
"A"
TAN LINE
1 5/8"
6 1/4"
2 3/4"
"A"
B
H
9"
36"
G
SUPPORT LUG C/S
8" I.D.
3/4"
6"
12"
C
2 3/4"
TAN. LINE
2 1/4"
2 1/4"
E
D
F
I
J
SPECIFICAT
DESIGN P
DESIGN T
MATERIA
MANUF
WITH ASME
PENNA. C




D Bottom
I Bottom
C
B
3"
1 3/4"
2"
1"
2 1/2"
1/2" Holes
A
Top
E
Bottom
F
Bottom
J
Bottom
G
H

~ ORIENTATION ~


1"
1 3/4 R
1/4" O.D. Tubing
2 1/2"
1"
J
I

SECTION "A-A"

NOZZLE SCHEDULE		
ITEM	DESCRIPTION	SERVICE
A	1/2"-300# W.N. FLG. R.F. & BLI	GAS INLET
B	1/4" COUPLING	GAS OUTLET
C	1/4" COUPLING	WATER INLET
D	1/4" COUPLING	DRAIN
E	2" COUPLING	HEATER
F	2" COUPLING	HEATER
G	1/2" HALF CPLG.	GAGE GLASS
H	1/2" HALF CPLG.	GAGE GLASS
I	1/4" O.D. TUBE	THERMOWELL
J	1/4" O.D. TUBE	THERMOWELL

FIGURE 20

SPECIFICATIONS:

DESIGN PRESSURE = 450 PSIG
DESIGN TEMPERATURE = 400 °F
MATERIAL: 304 S/S EXCEPT AS NOTED

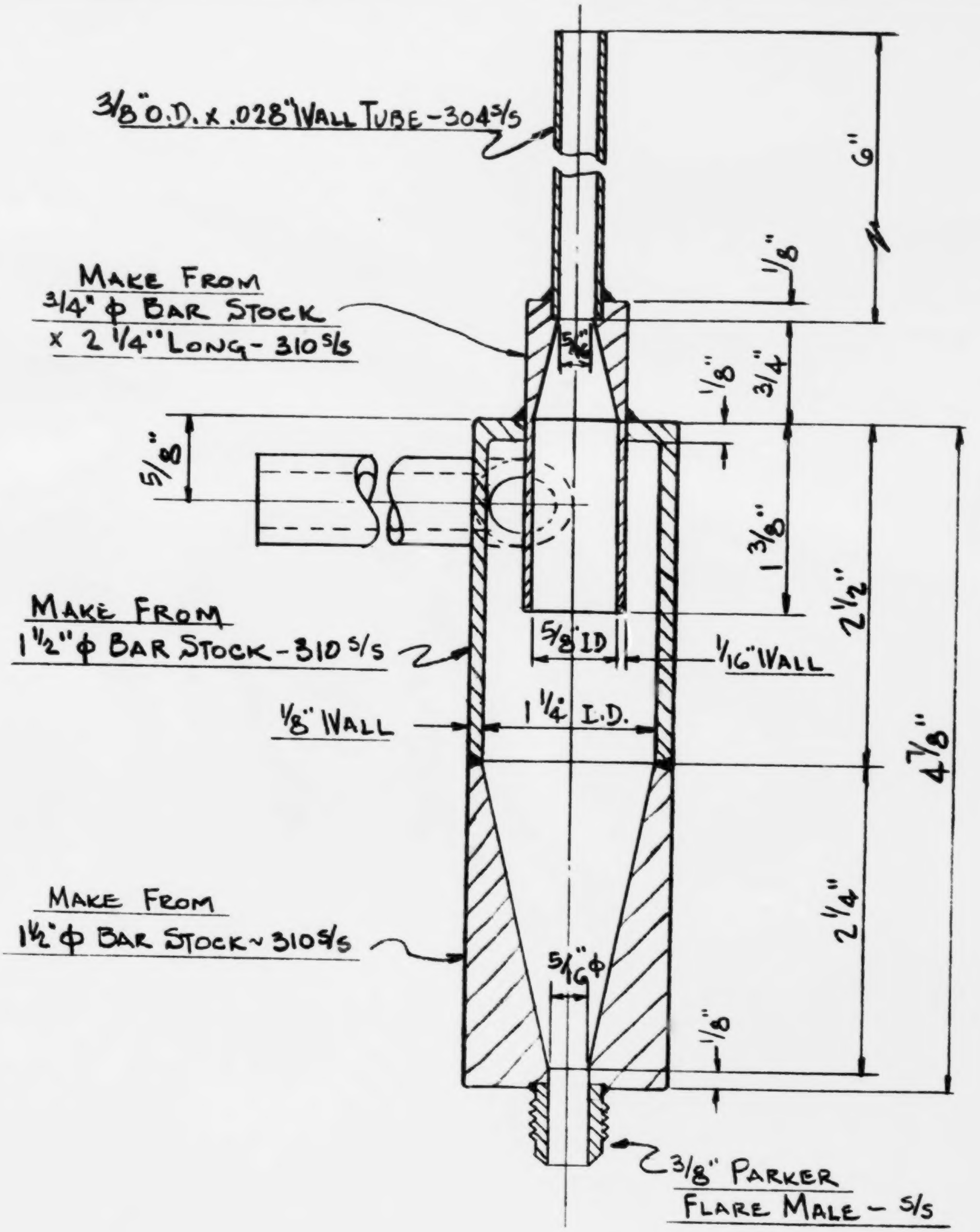
MANUFACTURED & STAMPED IN ACCORDANCE WITH ASME PRESSURE VESSEL CODE SECT. VIII
PENNA. CODE STAMP REQUIRED.

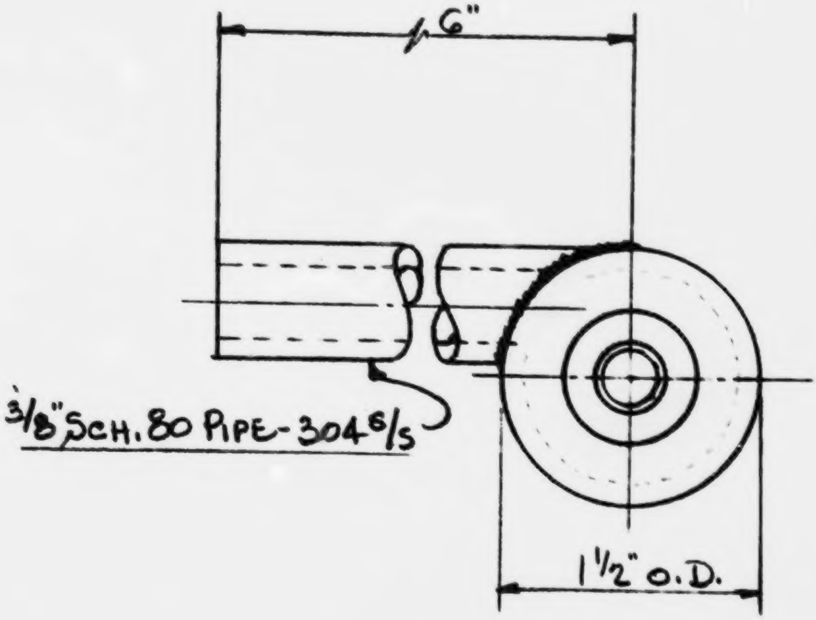
REVISIONS		
DRAWN	WAS	1/13/71
CHKD.		
APPR.		

CONSOLIDATION COAL CO.
RESEARCH DIVISION
LIBRARY, PENNA.

B-1 SATURATOR
CONTINUOUS CO_2 ACCEPTOR UNIT

SCALE	PROJECT	DRAWING NO.
3/8" = 1"	551.00	BV-3136


3/8" O.D. x .028" Wall Tube - 304 S/S
6"
1/8"
Make From
3/4" φ Bar Stock
x 2 1/4" Long - 310 S/S
5/16"
1/8"
3/4"
5/8"
1 3/8"
2 1/2"
4 7/8"
Make From
1 1/2" φ Bar Stock - 310 S/S
5/8" ID
1/16" Wall
1/8" Wall
1 1/4" I.D.
Make From
1 1/2" φ Bar Stock - 310 S/S
5/16" φ
2 1/4"
1/8"
3/8" Parker
Flare Male - S/S



~PLAN VIEW~

DESIGN CONDITIONS:
400 PSIG @ 900°F

HYDROSTATIC TEST : 800 PSIG
MATERIAL : AS NOTED

FIGURE 21

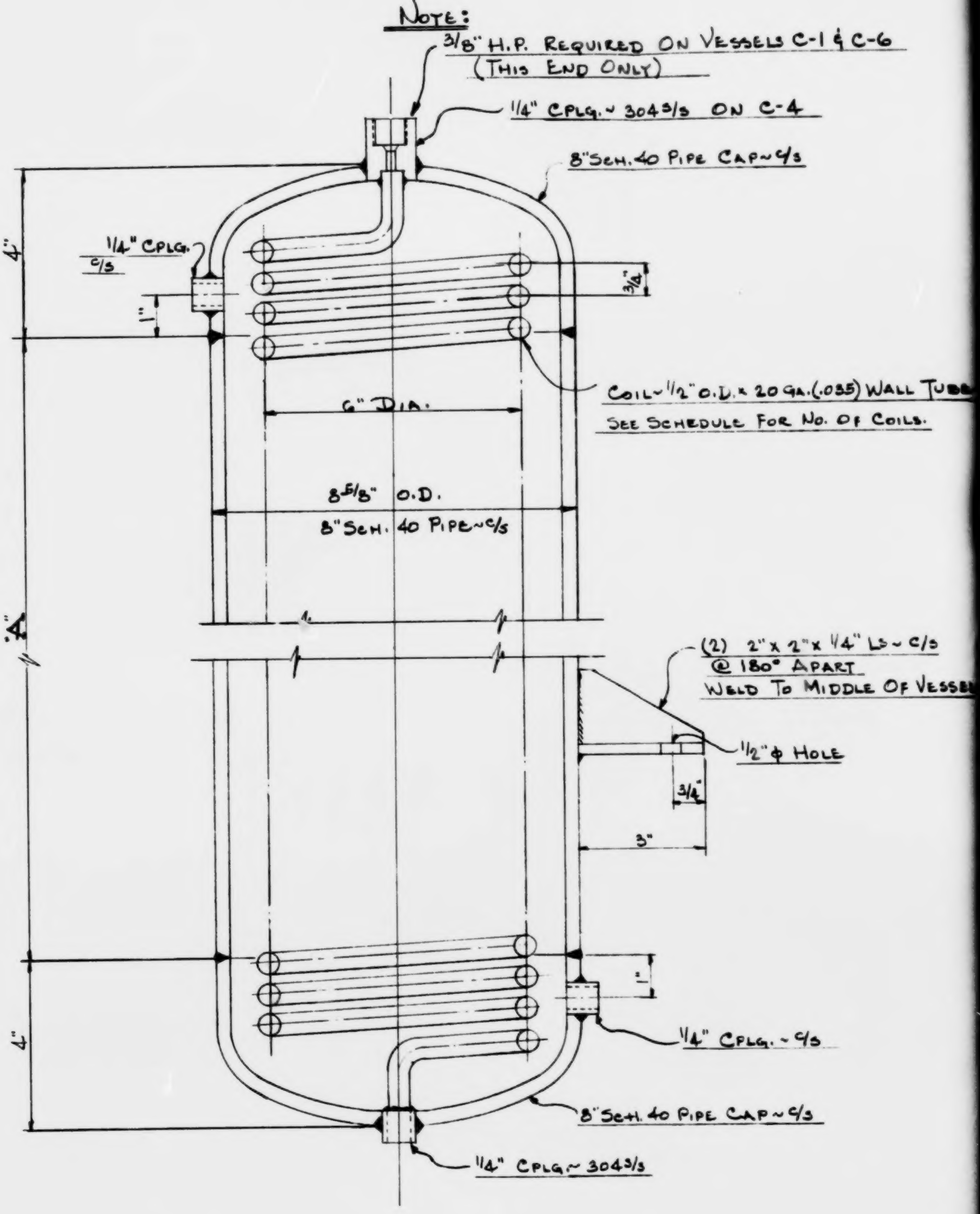
REVISIONS		

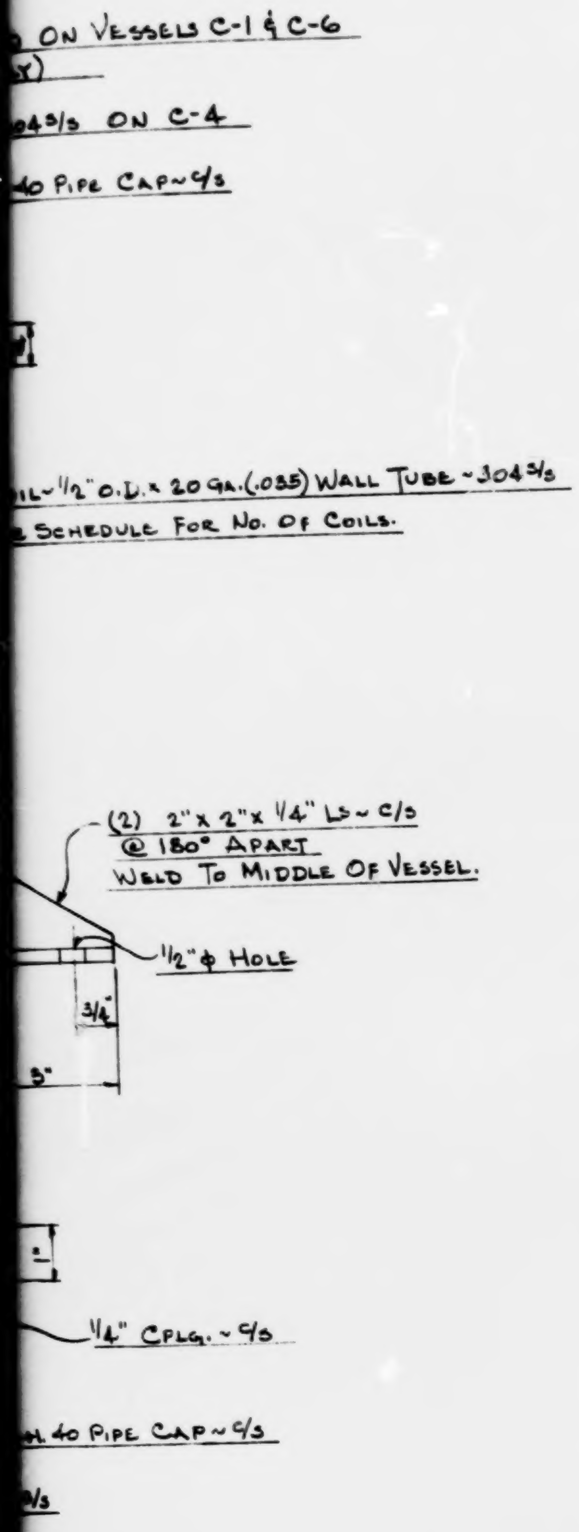
CONSOLIDATION COAL CO.
RESEARCH DIVISION
LIBRARY, PENNA.

G-1 CYCLONE FOR
CONTINUOUS CO_2 ACCEPTOR UNIT

DRAWN	WAS	12-16-70
CHKD.		
APPR.		

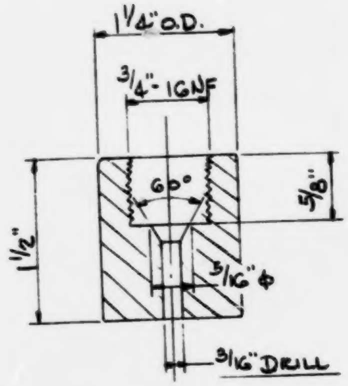
SCALE	PROJECT	DRAWING NO.
FULL	551.00	CV-3128


NOTE:
3/8" H.P. REQUIRED ON VESSELS C-1 & C-6
(THIS END ONLY)
1/4" CPLG. ~ 304 S/S ON C-4
8" SCH. 40 PIPE CAP ~ C/S
1/4" CPLG. C/S
4"
1"
3/4"
COIL ~ 1/2" O.D. x 20 GA. (.035) WALL TUB
SEE SCHEDULE FOR NO. OF COILS.
6" DIA.
8 5/8" O.D.
8" SCH. 40 PIPE ~ C/S
"A"
(2) 2" x 2" x 1/4" LS ~ C/S
@ 180° APART
WELD TO MIDDLE OF VESSE
1/2" ϕ HOLE
3/4"
3"
1"
4"
1/4" CPLG. ~ C/S
8" SCH. 40 PIPE CAP ~ C/S
1/4" CPLG. ~ 304 S/S


ON VESSELS C-1 & C-6
Y)
04 S/S ON C-4
40 PIPE CAP ~ C/S
IL ~ 1/2" O.D. × 20 GA. (.035) WALL TUBE ~ 304 S/S
SCHEDULE FOR NO. OF COILS.
(2) 2" × 2" × 1/4" LS ~ C/S
@ 180° APART
WELD TO MIDDLE OF VESSEL.
1/2" ϕ HOLE
3/4"
3"
1/4" CPLG. ~ C/S
H. 40 PIPE CAP ~ C/S
/S

~ COIL SCHEDULE ~

VESSEL No.	DIMENSION "A"	No. COILS	APPROX. FT. OF 1/2" TUBE REQ'D.
C-1	20"	32	50'
C-4	3"	9	14'
C-6	8"	16	25'


1 1/4" O.D.
3/4"-16NF
60°
5/8"
1 1/2"
3/16" ϕ
3/16" DRILL

DETAIL FOR 3/8" H.P.

MAT'L: 304 S/S 2 REQ'D.

-DESIGN CONDITIONS~

VESSEL	PSIG	TEMP.	HYDRO TEST
C-1	450	600°F	675 PSIG
C-4	450	100°F	675 PSIG
C-6	450	800°F	850 PSIG

MATERIAL : AS NOTED

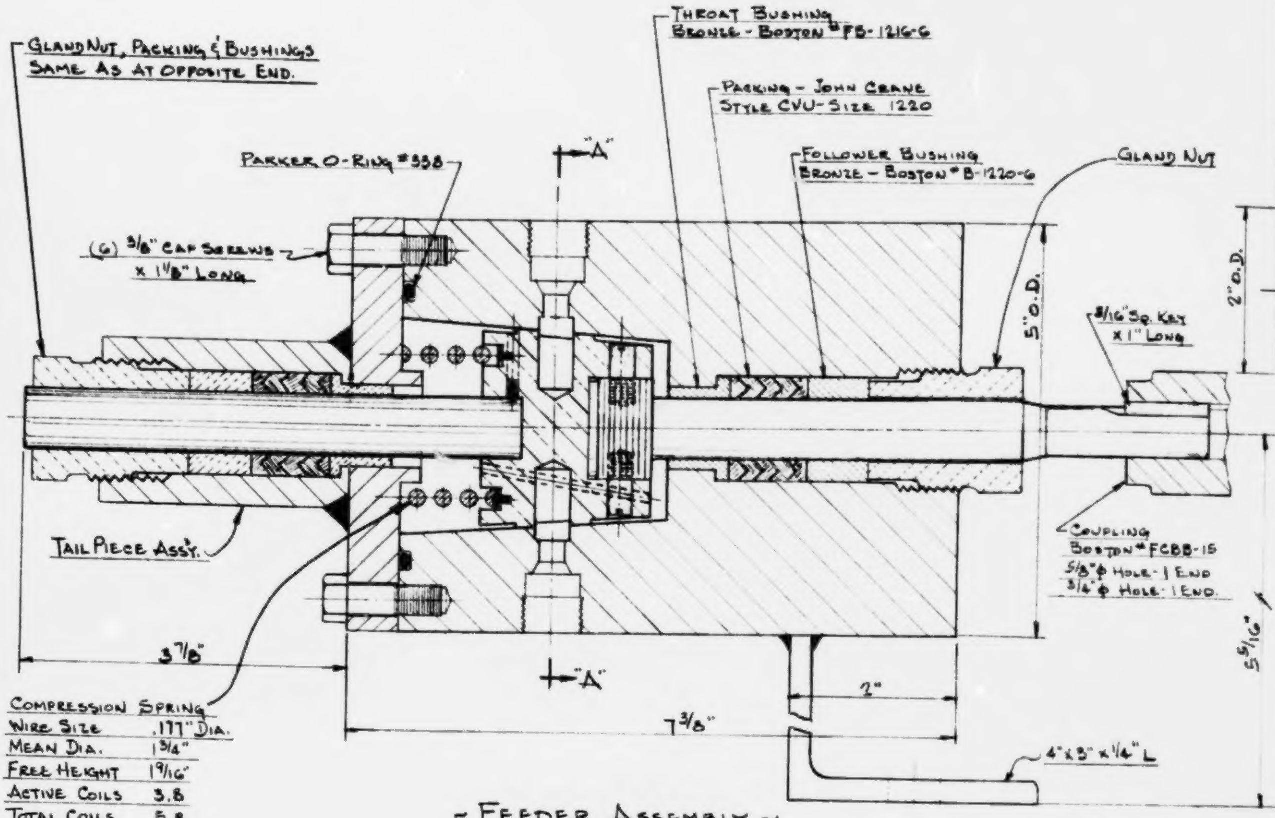
FIGURE 22

REVISIONS

CONSOLIDATION COAL CO.
RESEARCH DIVISION
LIBRARY, PENNA.

C-1, C-4, C-6 COOLERS
CONTINUOUS CO2 ACCEPTOR
GASIFICATION UNIT

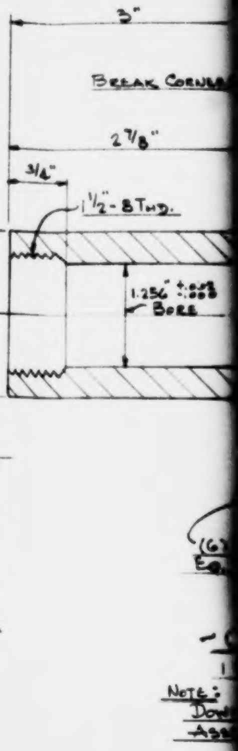
			SCALE	PROJECT	DRAWING NO.
DRAWN	WS	5/14/65	1/2" = 1"	531.03	BV-2696
CHKD.	P.F.M	5/19/65			
APPR.					

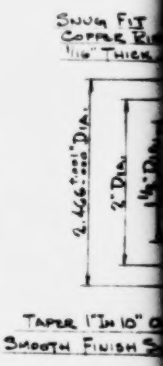

GLAND NUT, PACKING & BUSHINGS SAME AS AT OPPOSITE END.
THROAT BUSHING BRONZE - BOSTON # FB-1216-6
PACKING - JOHN CRANE STYLE CVU - SIZE 1220
FOLLOWER BUSHING BRONZE - BOSTON # B-1220-6
GLAND NUT
PARKER O-RING # 338
"A"
(6) 3/8" CAP SCREWS x 1 1/8" LONG
5" O.D.
3/16" SQ. KEY x 1" LONG
TAIL PIECE ASS'Y.
COUPLING BOSTON # FCBB-15 5/8" φ HOLE - 1 END 3/4" φ HOLE - 1 END.
3 7/8"
"A"
2"
7 3/8"
5 5/16"
4" x 3" x 1/4" L

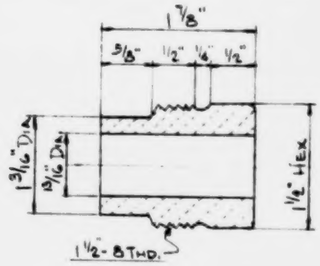
COMPRESSION SPRING	
WIRE SIZE	.177" DIA.
MEAN DIA.	1 3/4"
FREE HEIGHT	1 9/16"
ACTIVE COILS	3.8
TOTAL COILS	5.8
SCALE	66.7 #/IN.
MAT'L.	302 S/S
ENDS CLOSED & GROUND	

- FEEDER ASSEMBLY -

SCALE - FULL

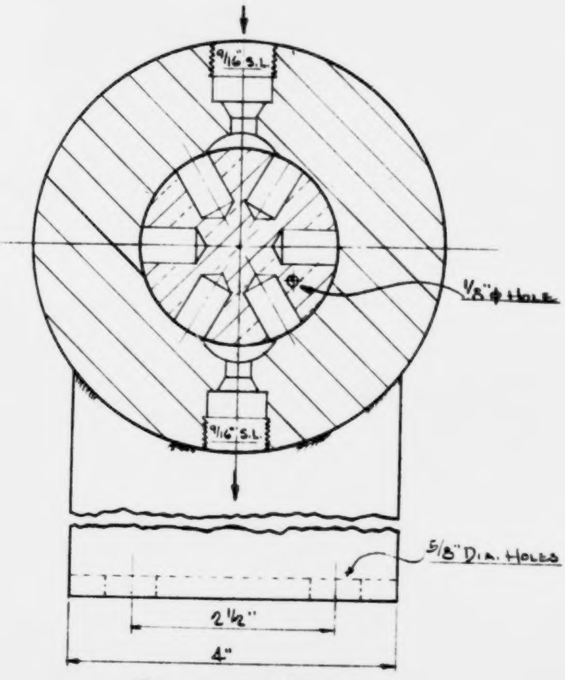

3"
BREAK CORNER
2 7/8"
3/4"
1 1/2 - 8 THD.
2" O.D.
1.256" BORE
(6)
NOTE:


SNUG FIT
COPPER R
1/16" THICK
2.466" DIA.
2" DIA.
TAPER 1" IN 10"
SMOOTH FINISH


1 7/8"
5/8"
1/2"
1/4"
1/2"
1 3/16" DIA.
13/16" DIA.
1 1/2" HEX
1 1/2 - 8 THD.

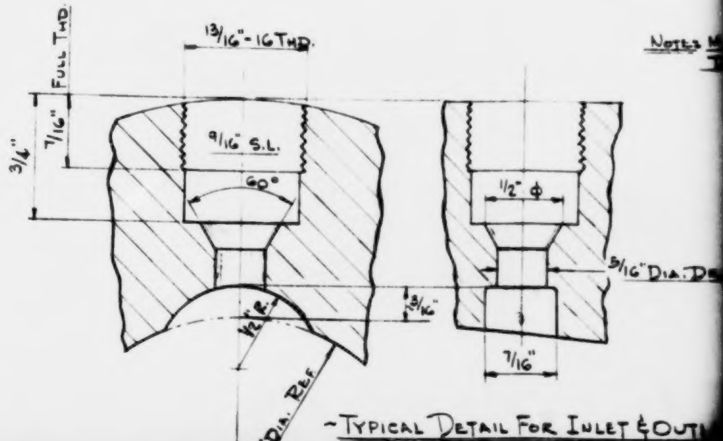
- PACKING GLAND NUT -

2 REQ'D. - MAT'L. BRONZE


9/16 S.L.
1/8" φ HOLE
9/16" S.L.
5/8" DIA. HOLES
2 1/2"
4"

- SECTION "A-A" -

SCALE - FULL


FULL THD.
13/16" - 16 THD.
3/4"
7/16"
9/16" S.L.
60°
3/16"
1/2" R
DIA. REF
1/2" φ
5/16" DIA. D
7/16"
NOTES

- TYPICAL DETAIL FOR INLET & OUT

SCALE - 2 TIMES

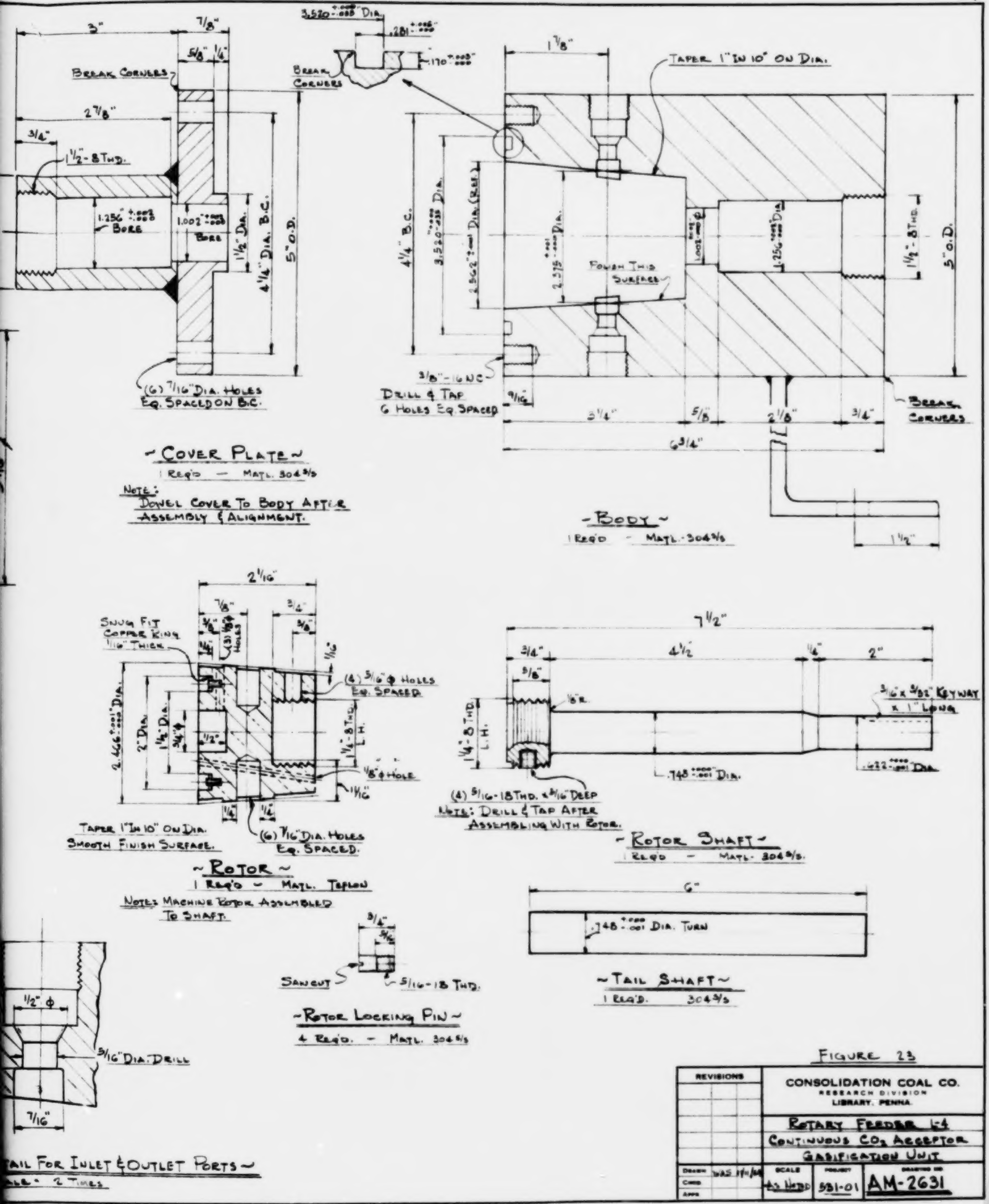
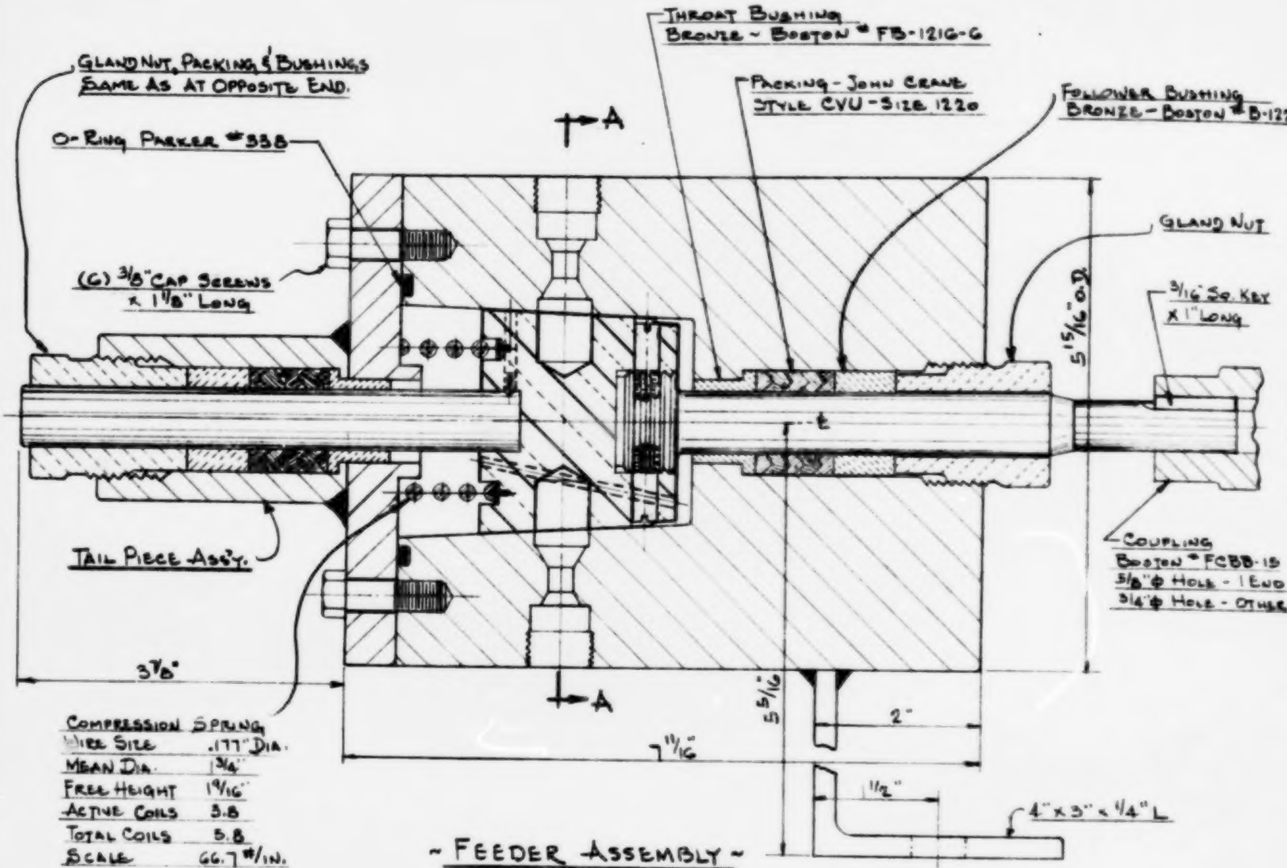

3"
7/8"
5/8"
1/4"
Break Corners
2 7/8"
3/4"
1 1/2"-8 Thd.
1.256" Bore
1.002" Bore
1 1/2" Dia.
4 1/4" Dia. B.C.
5" O.D.
(6) 7/16" Dia. Holes Eq. Spaced on B.C.
~ Cover Plate ~
1 Req'd — Matl. 304 S/S
Note: Dowel Cover To Body After Assembly & Alignment.
3.520 Dia.
.281
.170
Break Corners
1 7/8"
Taper 1" In 10" On Dia.
4 1/4" B.C.
3.520 Dia.
2.562 Dia. (Ref.)
2.375 Dia.
Finish This Surface
1.002 Dia.
1.256 Dia.
1 1/2"-8 Thd.
5" O.D.
3/8"-16 NC
Drill & Tap 6 Holes Eq. Spaced
9/16
3 1/4"
5/8"
2 1/8"
3/4"
Break Corners
6 3/4"
1 1/2"
- Body -
1 Req'd - Matl. 304 S/S
2 1/16"
7/8"
3/4"
3/8"
3/8"
(3) 1/8" Holes
1/16
1/2"
Snug Fit Copper Ring 1/16" Thick
(4) 5/16" Holes Eq. Spaced
2.466 Dia.
2" Dia.
1 1/2" Dia.
1 3/4"
1 1/4"-8 Thd. L.H.
1/8" Hole
1/16"
1/4"
1/4"
Taper 1" In 10" On Dia. Smooth Finish Surface.
(6) 7/16" Dia. Holes Eq. Spaced.
~ Rotor ~
1 Req'd ~ Matl. Teflon
Note: Machine Rotor Assembled To Shaft.
7 1/2"
3/4"
4 1/2"
1/4"
2"
5/8"
1/8 R.
3/16 x 3/32 Keyway x 1" Long
1 1/4"-8 Thd. L.H.
.748 Dia.
.422 Dia.
(4) 5/16-18 Thd. x 5/16 Deep
Note: Drill & Tap After Assembling With Rotor.
- Rotor Shaft -
1 Req'd - Matl. 304 S/S
6"
.748 Dia. Turn
~ Tail Shaft ~
1 Req'd. 304 S/S
3/4"
3/16
Saw cut
5/16-18 Thd.
- Rotor Locking Pin -
4 Req'd. - Matl. 304 S/S
1/2" Φ
3/16" Dia. Drill
7/16"
tail For Inlet & Outlet Ports ~
le - 2 Times

Figure 23

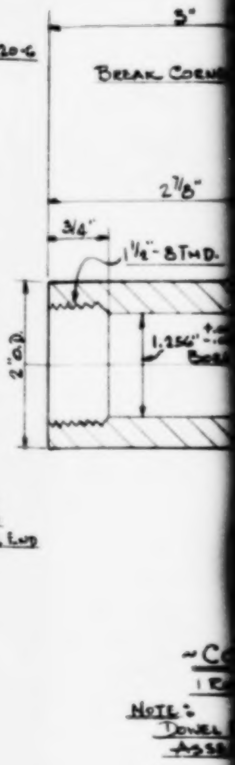
Revisions	CONSOLIDATION COAL CO. RESEARCH DIVISION LIBRARY, PENNA.
	Rotary Feeder L-4 Continuous CO₂ Acceptor Gasification Unit
Drawn WAS 1/1/64	Scale As Noted · Project 531-01 · Drawing No. AM-2631


Gland Nut, Packing & Bushings Same As At Opposite End.
O-Ring Parker #338
(6) 3/8" Cap Screws x 1 1/8" Long
Tail Piece Ass'y.
Throat Bushing Bronze ~ Boston # FB-1216-6
Packing - John Crane Style CVU - Size 1220
Follower Bushing Bronze - Boston # B-1220-6
Gland Nut
3/16" Sq. Key x 1" Long
5 15/16" O.D.
Coupling Boston # FCBB-15 5/8"φ Hole - 1 End 3/4"φ Hole - Other End
A
A
℄
3 7/8"
5 5/16
2"
7 11/16"
1 1/2"
4" x 3" x 1/4" L

Compression Spring	
Wire Size	.177" Dia.
Mean Dia.	1 3/4"
Free Height	1 9/16"
Active Coils	3.8
Total Coils	5.8
Scale	66.7 #/in.
Mat'l.	302 S/S
Ends Closed & Ground	

~ FEEDER ASSEMBLY ~

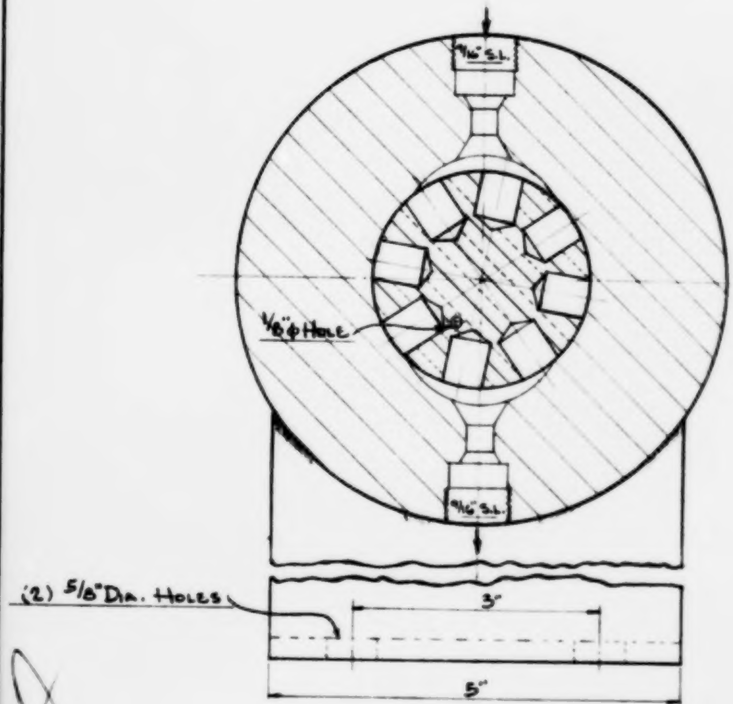
SCALE - FULL


3"
Break Corn
2 7/8"
3/4"
1 1/4"-8 THD.
2" O.D.
1.256"
Bore

~CO

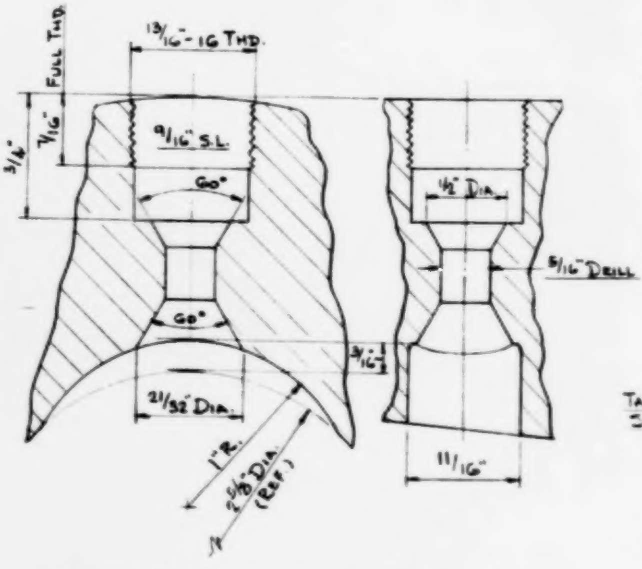
1 R

NOTE:
Dowel
Asse


9/16" S.L.
1/8"φ Hole
9/16" S.L.
(2) 5/8" Dia. Holes
3"
5"

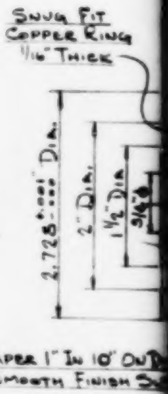
~SECTION "A-A"~

SCALE - FULL


Full Thd.
13/16"-16 THD.
3/16"
7/16"
9/16" S.L.
60°
60°
21/32" Dia.
1" R.
2 5/8" Dia. (Ref.)
1/2" Dia.
5/16" Drill
3/16"
11/16"

~TYPICAL DETAIL FOR INLET & OUTLET PORTS~

SCALE - 2 TIMES

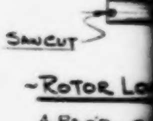

Snug Fit Copper Ring 1/16" Thick
2.728 +.001 -.000 Dia.
2" Dia.
1 1/2" Dia.
3/16"

Taper 1" In 10" On D
Smooth Finish S

~

1 R

Note: Machin



~Rotor Lo

4 Req'd. -

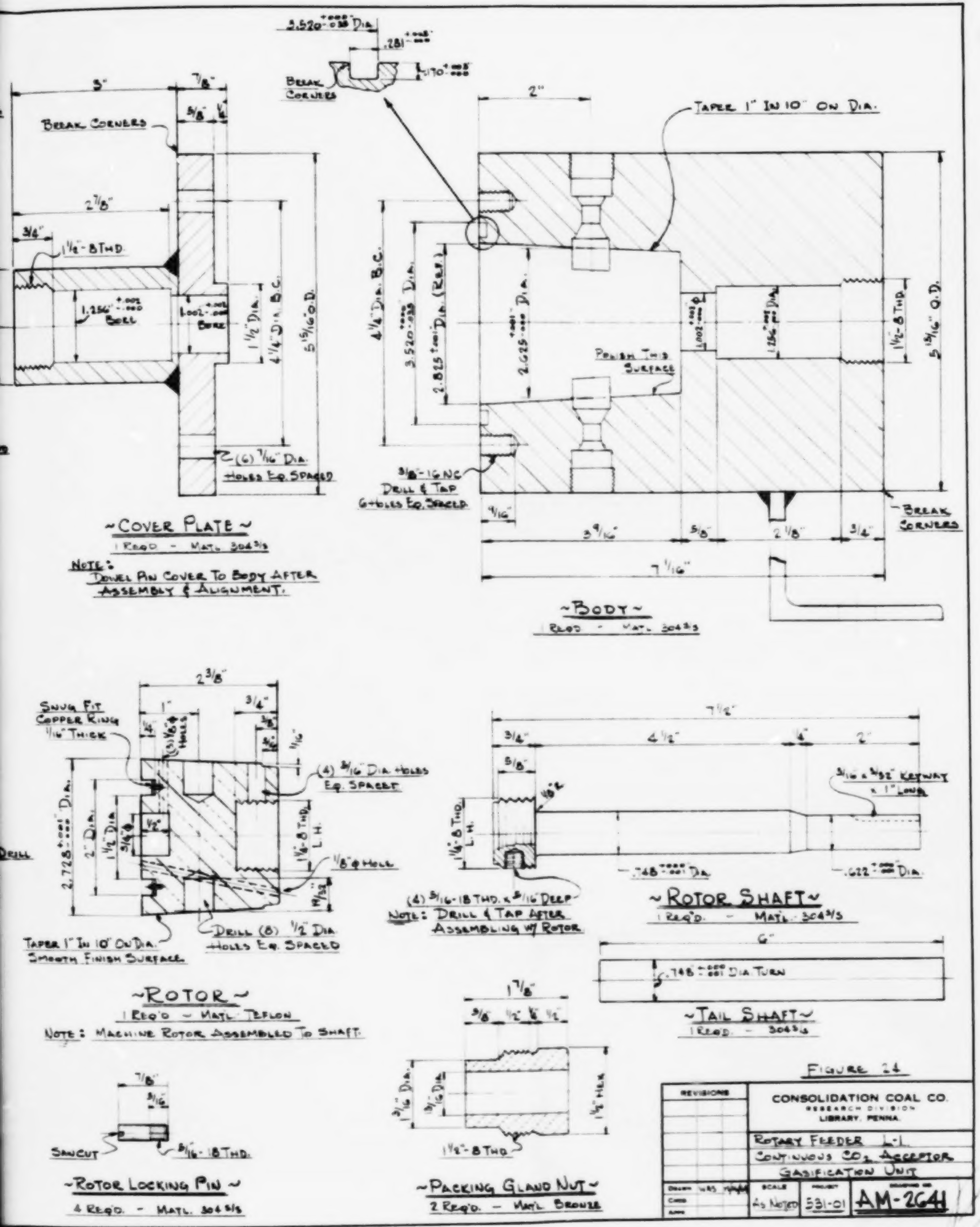

3.520 +.000 -.038 Dia.
.281 +.005 -.000
.170 +.005 -.000
Break Corners
3"
7/8"
5/8"
1/4"
Break Corners
2 7/8"
3/4"
1 1/2 - 8 THD.
1.256" +.002 -.000 Bore
1.002 +.002 -.000 Bore
1 1/2" Dia.
4 1/4" Dia. B.C.
5 15/16" O.D.
(6) 7/16" Dia. Holes Eq. Spaced
~Cover Plate~
1 Req'd - Mat'l. 304 S/S
Note: Dowel Pin Cover to Body After Assembly & Alignment.
2"
Taper 1" in 10" on Dia.
4 1/4" Dia. B.C.
3.520 +.000 -.038 Dia.
2.825 +.001 Dia. (Ref.)
2.625 +.001 -.000 Dia.
Polish This Surface
1.002 +.002 -.000 D
1.256" +.002 Dia
1 1/2 - 8 THD.
5 15/16" O.D.
3/8"-16 NC Drill & Tap 6 Holes Eq. Spaced
9/16"
3 9/16"
5/8"
2 1/8"
3/4"
Break Corners
7 1/16"
~Body~
1 Req'd - Mat'l. 304 S/S
2 3/8"
1"
3/4"
3/8"
3/16"
1/16"
1/4"
(3) 1/8" ϕ Holes
Snug Fit Copper Ring 1/16" Thick
(4) 5/16" Dia. Holes Eq. Spaced
2.728 +.001 -.000 Dia.
2" Dia.
1 1/2" Dia
3/4" ϕ
1/2"
1 1/2 - 8 THD. L.H.
1/8" ϕ Hole
11/32
Taper 1" in 10" on Dia. Smooth Finish Surface
Drill (8) 1/2" Dia Holes Eq. Spaced
~Rotor~
1 Req'd - Mat'l. Teflon
Note: Machine Rotor Assembled to Shaft.
7 1/2"
3/4"
4 1/2"
1/4"
2"
5/8"
1/8"
1 1/2 - 8 THD. L.H.
3/16 x 3/32" Keyway x 1" Long
.748 +.000 -.001 Dia.
.622 +.000 -.001 Dia.
(4) 5/16-18 THD. x 5/16" Deep
Note: Drill & Tap After Assembling w/ Rotor.
~Rotor Shaft~
1 Req'd - Mat'l. 304 S/S
6"
.748 +.000 -.001 Dia. Turn
~Tail Shaft~
1 Req'd - 304 S/S
7/8"
3/16"
Sawcut
5/16 - 18 THD.
~Rotor Locking Pin~
4 Req'd - Mat'l. 304 S/S
1 7/8"
3/8
1/2
1/8
1/2
3 1/16" Dia.
3/16" Dia.
1 1/2" - 8 THD.
1 3/4" Hex
~Packing Gland Nut~
2 Req'd - Mat'l. Bronze

Figure 24

Revisions

CONSOLIDATION COAL CO.
RESEARCH DIVISION
LIBRARY, PENNA.

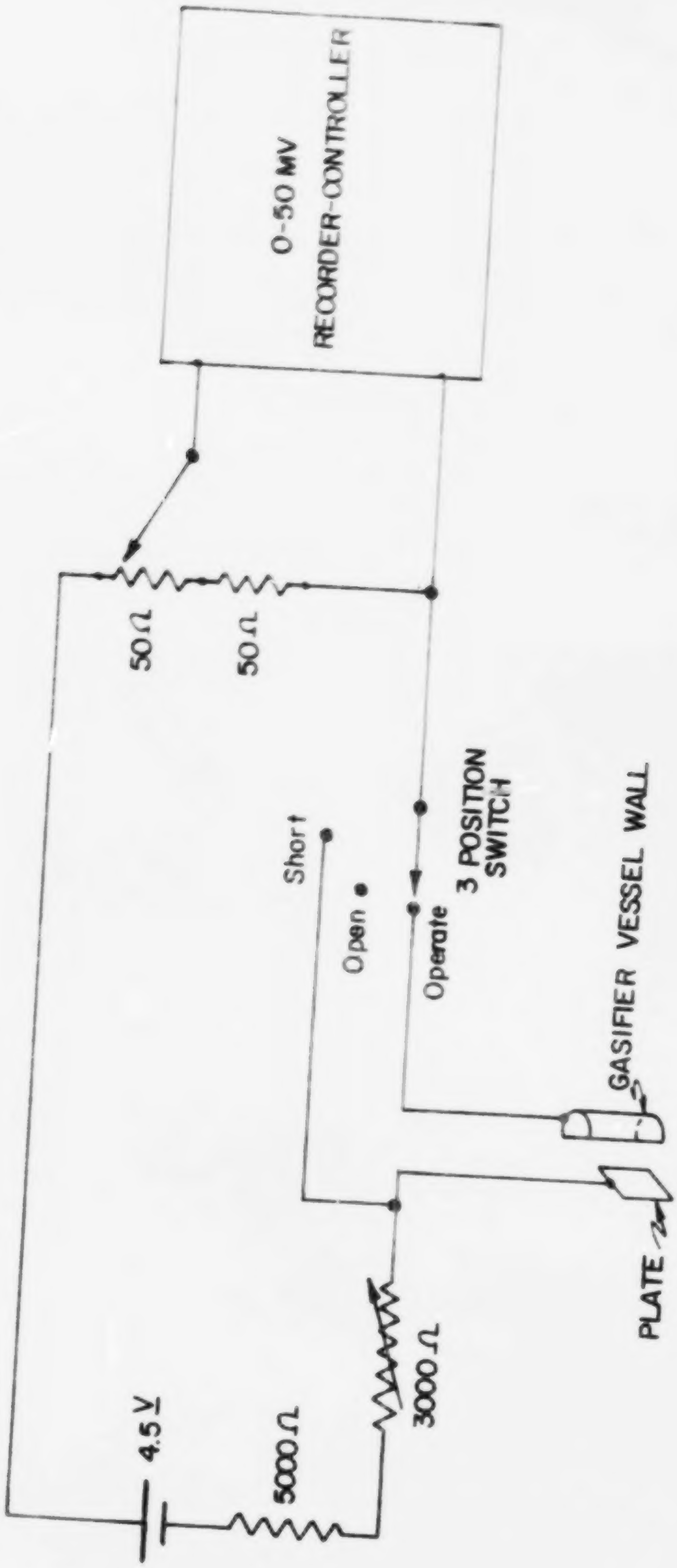
Rotary Feeder L-1
Continuous CO_2 Acceptor
Gasification Unit

Drawn WRS	Scale: As Noted	Project: 531-01	Drawing No.: AM-2641
Chk'd			
App'd			

Figure 25

SCHEMATIC DIAGRAM

CHAR-ACCEPTOR INTERFACE PROBE CIRCUIT


0-50 MV
RECORDER-CONTROLLER
50 Ω
50 Ω
3 POSITION SWITCH
Short
Open
Operate
GASIFIER VESSEL WALL
PLATE
3000 Ω
5000 Ω
4.5 V

B. Materials

For the convenience of the reader, properties of the feedstocks used during the experimental program are summarized here. This information also is introduced at the appropriate point in Section V, Results and Discussion.

1. Preoxidizer Feeds

Two types of bituminous coals were fed to the preoxidizer, Illinois No. 6 and Pittsburgh seam. The Pittsburgh seam coal was from Consol's Ireland Mine and the Illinois No. 6 coal was from Consol's Hillsboro Mine. Both are high-sulfur, high-volatile steam coals. The Ireland Mine coal is among the world's most highly fluid coals when heated through the temperature range of 700-900°F. The composition and size consist of the coal feedstocks which were fed to the gasifier are given in Table A.

2. Gasifier Feeds

Gasifier operation was first tested with a feedstock which was known to be inert with respect to agglomeration. For this purpose, Disco bituminous coal char from Consol's reference stock kept at Library, Pa. was used. Disco char is an LTC smokeless fuel formerly produced by Consol. Disco was made from Pittsburgh seam coal and was suitable for testing the possibility of ash slagging in the gasifier since its ash composition is typical of Pittsburgh seam coals. Disco was ground and sized to 28 x 100 Tyler mesh and then was devolatilized in nitrogen at 1000°F. Table B gives data on the properties of this feed.

The first four gasification runs were made with pregasified (at 1775°F) Disco char produced in a previous experimental program which was privately sponsored. The composition of this material also is shown in Table B.

Preoxidized coal also was fed to the gasifier. The properties of those preoxidized coals which were fed successfully are also given in Table B. A complete discussion of the preoxidation conditions which produced these feeds is given in Section V.

3. Regenerator Feeds

The acceptor fed to the regenerator was Tymochtee dolomite obtained from C. E. Duff and Son, Inc., Huntsville, Ohio. The chemical analysis of this acceptor is given in Table C.

The fuel char fed to the regenerator was Disco char which had been fed through the gasifier at 1700°F during break-in operations prior to Run D1B. The analysis of this material is given in Table D.

TABLE A

Properties of Preoxidizer Feedstocks

Material	Pittsburgh Seam Ireland Mine Feed Coal	Illinois No. 6 Coal Hillsboro Mine Feed Coal (1)	Illinois No. 6 Coal Hillsboro Mine Feed Coal (2)
Hydrogen, Wt % (dry basis)	5.00	4.19	4.18
Carbon	72.24	60.44	60.14
Nitrogen	1.35	1.17	1.22
Oxygen (by difference)	5.53	8.81	7.67
Sulfur	4.52	4.46	4.93
Ash	11.36	20.93	21.86
Volatile Matter	40.1	37.2	36.4
Pyritic Sulfur	2.53	2.20	2.26
Sulfate Sulfur	.07	.07	.03
Size, Tyler Mesh, Wt %			
-24	0	0	0
24 x 28	11.2	9.6	15.8
28 x 35	35.4	39.3	41.1
35 x 48	34.8	34.2	29.0
48 x 65	16.9	15.0	12.7
65 x 100	1.2	1.3	1.0
-100	0.5	0.6	0.4
Mean Diameter, in.(3)	.0165	.0166	.0176
Mean Density, lb/ft^3(4)	81.0	80.0	80.0

(1) Batch No. 1 used in Runs 5P1 and 5P2.
(2) Batch No. 2 used in Run 6P.
(3) Arithmetic mean.
(4) Measured in mercury at 1 atmosphere.

TABLE B

Properties of Gasifier Feedstocks

Feed Material	Disco Char	Pregasified Disco Char	Run 4P	Run 5P2	Run 6P
			← Preoxidized Coal		→
Used in Run Number	D1B	OG2-OG5	4G3	5G3	6G3
Hydrogen, Wt % (dry basis)	2.63	0.30	3.26	3.29	3.44
Carbon	76.24	78.10	69.81	67.02	70.61
Nitrogen	1.77	0.64	1.43	1.44	1.36
Oxygen (difference)	4.66	.0	7.99	7.98	5.80
Sulfur	1.27	0.29	3.75	3.72	3.40
Ash	13.43	20.67	13.76	16.55	15.39
Volatile Matter			24.7		
Sulfide Sulfur	0		-	.01	.03
Pyritic Sulfur			1.57	1.11	1.37
Sulfate Sulfur			.11	.27	.07
Mean Diameter, in.(1)			.0156	.0171	.0175
Mean Density, lb/ft^3(2)			56.4	49.8	51.8
Ash Composition, Sulfur-Free Basis				(3)	
Al_2O_3, Wt %	23.2			18.5	
SiO_2	51.6			53.6	
Fe_2O_3	13.1			17.4	
CaO	4.7			3.7	(3,4)
MgO	0.8			1.1	
Na_2O	0.4			1.2	
K_2O	1.8			2.4	
TiO_2	1.0			0.9	
P_2O_5	0.4			0.3	
Total	97.0			99.1	

(1) Arithmetic mean.
(2) Measured in mercury at 1 atmosphere.
(3) Some segregation of discrete ash particles occurred during preoxidation.
(4) Approximately the same as 5P2.

TABLE C

Chemical Analysis of Tymochtee Dolomite

Al_2O_3, Wt %	0.91
SiO_2	5.21
Fe_2O_3	1.20
CaO	52.54
MgO	37.77
Na_2O	0.24
K_2O	0.35
TiO_2	0.70
P_2O_5	0.22
SrO	0.02
SO_3	0.85

TABLE D

Analysis of Regenerator Fuel Char

Hydrogen, Wt % (dry basis)	0.69
Carbon	76.98
Nitrogen	0.91
Oxygen (by difference)	1.72
Sulfur	0.84
Ash	18.86
Sulfide Sulfur	0.09

C. Feedstock Preparation

1. Feed Coal Preparation

The stepwise procedure used to prepare feed coal is shown in Figure 26. The coal was first dried at 200°F in a forced air circulation oven. This removed gross moisture in the coal, but when exposed to air, the coal quickly picked up moisture to some equilibrium level which is a function of the coal, temperature, and relative humidity.

The dry coal then was crushed in a hammermill. The crushed material was processed through a two deck vibrating screen at 24 and 100 Tyler mesh. The +24 mesh material was recycled to the hammermill and the -100 mesh was discarded. The remaining 24 x 100 mesh coal was used as feedstock.

Another feed material used was Disco char. This char was produced via low temperature carbonization of Pittsburgh seam coal and is nonagglomerating at gasifier and preoxidizer conditions. It was thus an ideal material for start-up operations. It was available in a 28 x 100 mesh size consist.

2. Fuel Char Preparation

Regenerator fuel char was gasifier product char from previous runs.

3. Acceptor Sizing

The kinetics of the acceptor recarbonation reaction dictated that the retention time of the acceptor showering through the gasifier char bed must be about seven minutes. The corresponding acceptor size was in the range of 16 x 48 Tyler mesh. The unit was operated using 16 x 28 mesh acceptor, in order to allow separation of a mixture of char and acceptor by screening, in case a sharp separation at the char-acceptor interface at the bottom of the gasifier did not occur.

The acceptor was ground using a coffee mill which was adjusted to maximize the yield of the desired 16 x 28 mesh size fraction. The ground acceptor was screened using a Ro-Tap at 16 x 28 Tyler mesh. The plus 16 mesh material was reground in the coffee mill and the minus 28 mesh fraction was set aside. The 16 x 28 mesh fraction was wet screened at 28 mesh to remove all of the fines. Wet screening was required to make attrition data obtained during operation of the unit meaningful.

After the wet-screened acceptor was dried under infrared lamps, the 16 x 28 mesh size was separated into 16 x 20 mesh and 20 x 28 mesh size fractions. Acceptor preparation is shown schematically in Figure 27.

Figure 26

FEED COAL PREPARATION

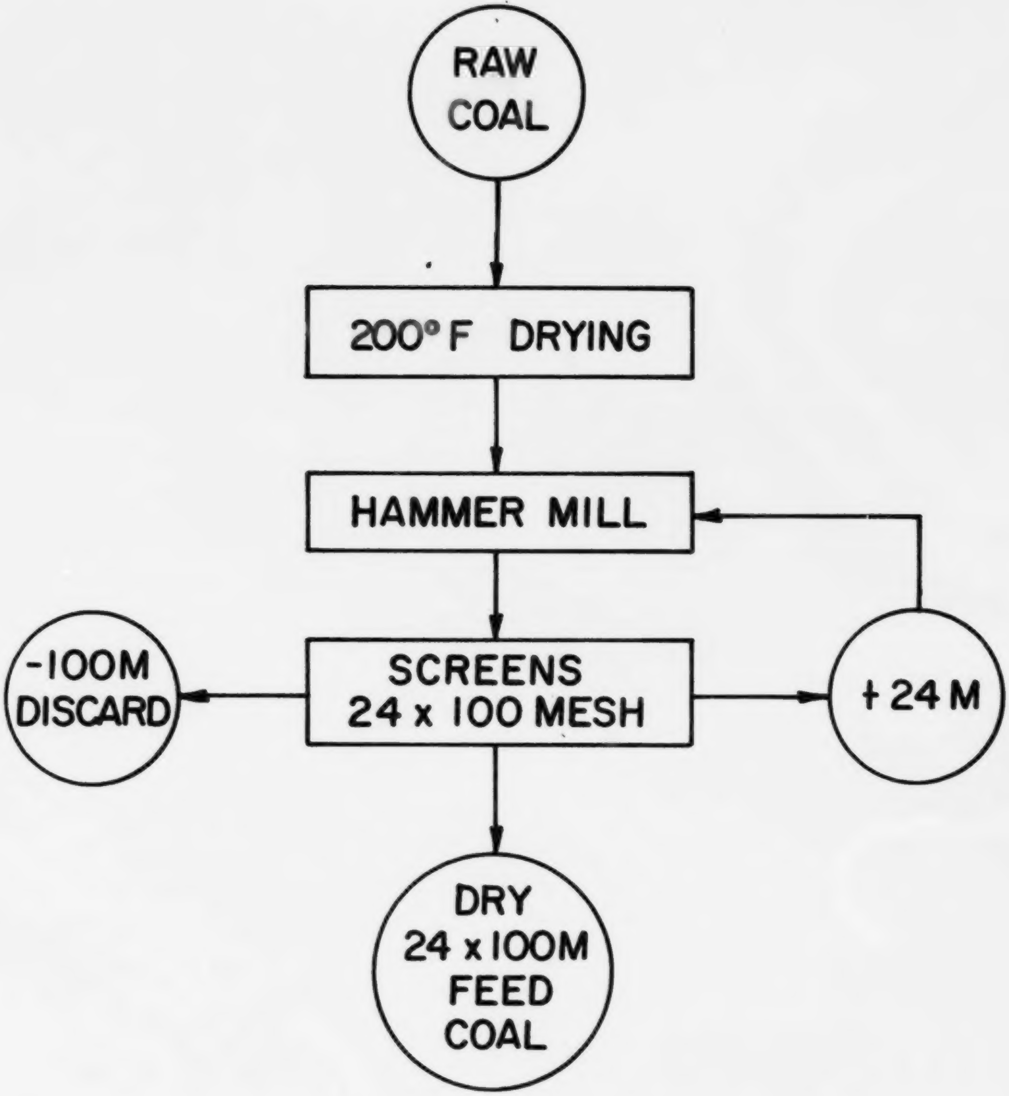
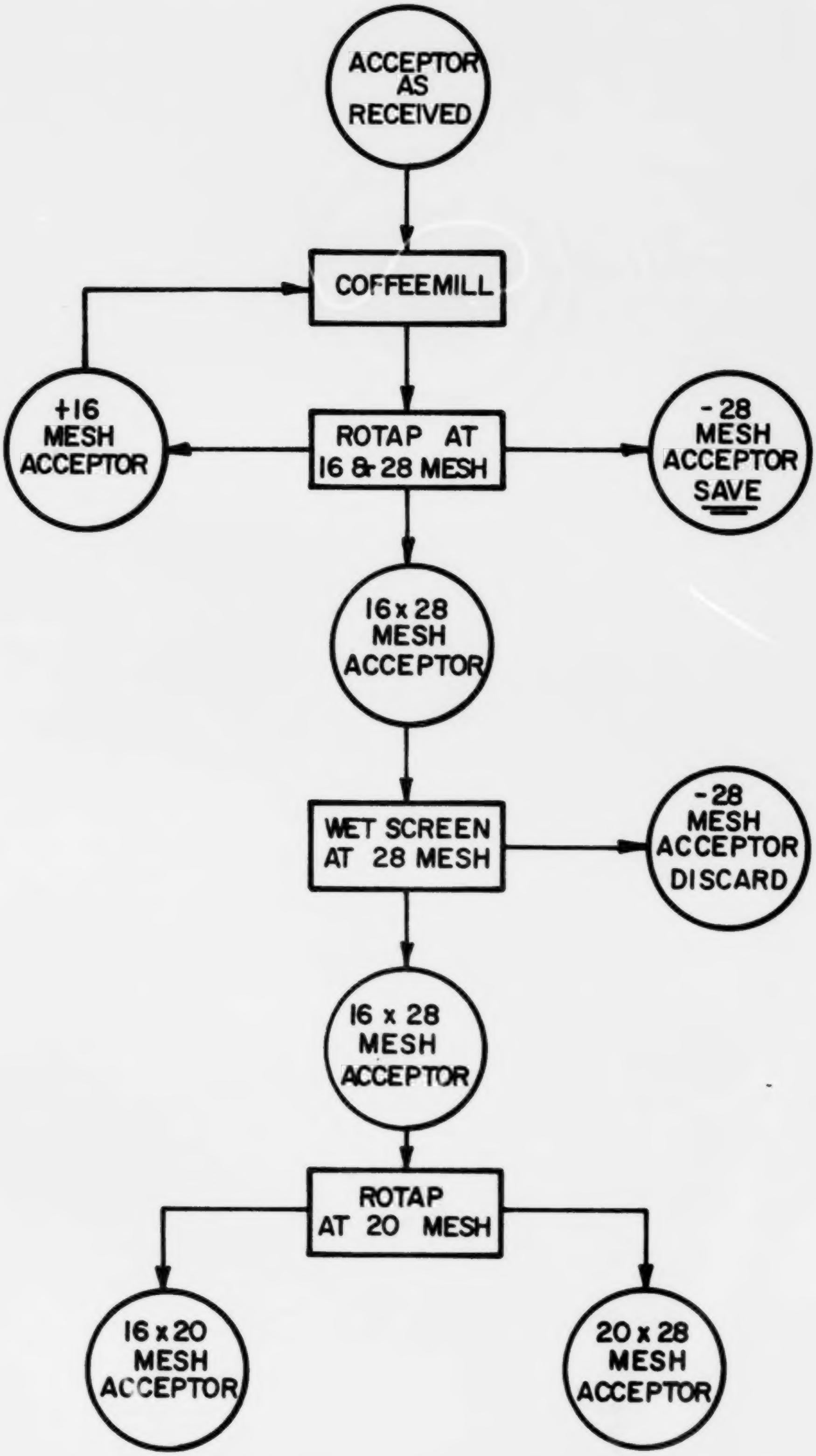

RAW COAL
200° F DRYING
HAMMER MILL
SCREENS 24 x 100 MESH
-100M DISCARD
+24 M
DRY 24 x 100M FEED COAL

Figure 27

ACCEPTOR SIZING


ACCEPTOR AS RECEIVED
COFFEEMILL
+16 MESH ACCEPTOR
ROTAP AT 16 & 28 MESH
-28 MESH ACCEPTOR SAVE
16 x 28 MESH ACCEPTOR
WET SCREEN AT 28 MESH
-28 MESH ACCEPTOR DISCARD
16 x 28 MESH ACCEPTOR
ROTAP AT 20 MESH
16 x 20 MESH ACCEPTOR
20 x 28 MESH ACCEPTOR

D. Operating Procedure for a Typical Two-Vessel Integrated Run

1. Prerun Procedure

Prior to starting a run, the feed hoppers were charged, the acceptor inventory was prepared, and the unit was leak tested.

A batch of pregasified char was charged to one of the gasifier feed hoppers to make up the initial gasifier char bed inventory, and a batch of from 5 to 50 pounds of the programmed gasifier feedstock was charged to the other feed hopper.

A batch of fuel char was charged to the fuel char feed hopper. Since the run would have to be stopped in order to recharge the hopper, the amount charged corresponded to either the hopper capacity or the amount of fuel char available, usually between 20 and 40 pounds.

The acceptor inventory was made up to the desired size consist by thoroughly mixing carefully weighed proportions of 16 x 20 mesh and 20 x 28 mesh size fractions. About 11,000 grams of raw acceptor were used in a typical run.

After bringing both the gasifier and regenerator to 1500°F via the electrical heaters, the unit was pressured up with N_2 to the programmed run pressure and was leak tested.

2. Startup

With the unit at pressure, the recycle compressors were started and, after setting the desired compressor discharge pressures, the programmed recycle flows to both vessels were established.

The steam flow was then put into the gasifier by diverting the gasifier fluidizing gas through the steam generator. The Powerstats on the steam generator heater were adjusted to control the programmed water temperature. The programmed flows of H_2 and CO_2 were established immediately after the steam flow. Early experience had shown that operation of the unit with H_2 and CO_2 in the absence of steam resulted in deposition of carbon in the gasifier preheater coil.

To insure a stable gasifier char-acceptor interface, the acceptor level was established in the gasifier prior to feeding char by taking 1,000 grams from the acceptor inventory and charging it to the gasifier through the acceptor charge pot.

The gasifier char feed was started from the feed hopper containing the pregasified char. The feed rate was about 10 lb/hr during this period. Char feed was continued until the gasifier char bed was established, and char began overflowing the weir. Then, the feeder drive setting was reduced to the rpm corresponding to the programmed char feed rate, and the Powerstats controlling the electrical heat input to the gasifier heater were adjusted to obtain the programmed gasifier bed temperature.

Feed was started from the char feed hopper containing the programmed run char feedstock when the hopper containing the pregasified char was empty, as shown by a decrease in the pressure drop across the char transfer line. The empty hopper was isolated from the system and was recharged with a batch of the programmed feedstock.

Next, the entire acceptor inventory was circulated through the unit in order to calcine the $MgCO_3$ component of the dolomite and to adjust the L-5 standleg and gasifier char-acceptor interface level controllers. During this period, the gasifier was operated at programmed conditions and the regenerator was held at 1600°F bed temperature and the CO_2 and N_2 flows were adjusted to give about 1.5 atm CO_2 partial pressure. At these conditions the $CaCO_3$ component of the acceptor will not calcine.

During this period, five charges of raw acceptor weighing 2,000 grams each were fed alternately from the acceptor feed hoppers. The L-4 feeder was set at 3 rpm which corresponded to a feed rate of 6 lb/hr (raw acceptor basis).

The emergency solenoid valve which isolated the process vessels was kept closed until the acceptor overflowing the regenerator weir filled the L-5 standleg.

As the acceptor was circulated, the L-5 valve positions corresponding to normal and low acceptor levels in the L-5 standleg and the L-6 valve setting corresponding to high, normal and low levels of the char-acceptor interface were adjusted until the optimum settings were obtained.

When an acceptor feed hopper became empty, the parallel hopper was switched onstream. Simultaneously, the acceptor withdrawal hoppers were switched. The off-stream withdrawal hopper was isolated and emptied. After all of the raw acceptor had been fed to the unit, the acceptor which had been removed from withdrawal hoppers was combined. Two batches of the combined acceptor weighing 2,300 grams each were made up and charged one batch to each feed hopper. These two batches constituted the initial external acceptor inventory for the run.

When all the raw acceptor had been fed, the L-4 feeder was shut off, and the regenerator acceptor bed was calcined by removing the CO_2 flow from the regenerator and raising the bed temperature to 1800°F. Calcining of the $CaCO_3$ component occurred in the temperature range of 1700 to 1750°F. The vessel wall temperatures were kept below 1950°F during calcining to eliminate the possibility of the acceptor agglomerating. Early experience had shown that mixtures of CaO and $CaCO_3$, rich in $CaCO_3$, would sinter at 1950°F.

After the acceptor feed hoppers were charged and the regenerator bed was calcined, the L-3 feed rpm was set to give the desired fuel char feed rate. The fuel char feed was started by opening the valve below the feed hopper. After waiting one minute to allow an inventory of fuel char to be established in the regenerator bed, the air flow was initiated and was slowly increased to the programmed rotameter setting. When the combustion reaction began, as shown by the rapid increase in the regenerator bed temperature, acceptor circulation was begun so that the sensible heat of the acceptor and the endothermic heat of the calcining reaction could act as a heat sink for the heat released by the combustion reaction. The start of acceptor circulation marked the beginning of the run.

3. Establishing Programmed Conditions

When the programmed air rotameter setting was reached, the Powerstats controlling the electrical heat input to the regenerator were adjusted to give the desired bed temperature.

As acceptor began showering down through the gasifier bed, additional CO_2 was added to the gasifier inlet gas to compensate for CO_2 which reacted with the calcined acceptor. The gasifier Powerstats were adjusted to compensate for the exothermic heat of the acceptor recarbonation reaction.

As soon as the acceptor circulation and bed temperatures were under control, a check was made to determine the CO and CO_2 concentration in the regenerator offgas. On the basis of these analyses, the air flow rate and fuel char feed rate were adjusted until the desired CO concentration was achieved. The flows of CO_2 and N_2 added to the regenerator inlet gas were adjusted to give the programmed reading on the CO_2 analyzer.

After the desired regenerator gas composition was attained, the gasifier product gas was analyzed using the gas chromatograph and the inlet flows of H_2 and CO_2 were adjusted to achieve the desired inlet gas composition.

Axial temperature traverses were taken across both fluidized beds as soon as the programmed bed temperatures were achieved. The relative electrical heat inputs to the heating zones were adjusted to give a flat temperature profile.

4. Routine Run Procedure

Continuous acceptor circulation was maintained by feeding acceptor to the regenerator alternately from the two acceptor feed hoppers. The acceptor feed hoppers and gasifier withdrawal hoppers were switched simultaneously. The off-stream withdrawal hopper was isolated from the unit and drained. The acceptor removed from the withdrawal hoppers was screened to plus 28 mesh and was charged to the off-stream acceptor feed hopper. Riffled 20-gram samples of the plus 28 mesh acceptor were taken at specified times during the run.

Gasifier char feed was maintained by feeding batches of the programmed run feedstock alternately from the two gasifier char feed hoppers.

The gasifier product char hopper was emptied at specified intervals which varied with the feed rate. The product char was screened at 28 mesh to remove any acceptor which came over the weir. The plus 28 mesh fraction was returned to the circulating acceptor inventory by adding it to an acceptor feed hopper charge. The minus 28 mesh fraction was weighed and then combined with the product char previously removed from the unit.

The ash receiver which collected the regenerator solids overhead material was emptied hourly. The overhead solids were screened at 35 mesh. The plus 35 mesh fraction, if any, was acceptor which was carried over during pressure upsets and was returned to the unit by adding it to an acceptor feed hopper charge. The minus 35 mesh material, which was essentially all residual fuel char and ash, was weighed and saved for analysis.

The flask which was used to collect the condensate drained from the condensate receiver was replaced hourly, and the weight of the condensate determined. The condensates collected in the sight glasses below the C-4 and C-6 coolers were drained and weighed at least once per shift.

The solids filters in the gasifier and regenerator outlet gas piping were changed when the ΔP across the on-stream filter was 75" H_2O. After the filters were switched, the off-stream filter was removed from the unit, cleaned, and reinstalled. The contents of each filter were weighed and saved for analysis.

As the acceptor lost activity during the course of the run, the amount of CO_2 picked up in the gasifier as the acceptor recarbonated, and liberated in the regenerator as the acceptor calcined, decreased. Periodic analyses of the gasifier product gas were made using the gas chromatograph. The concentration of CO_2 was monitored continuously on the IR analyzer. On the basis of these analyses, the CO_2 added to the gasifier inlet gas was decreased to maintain the desired CO_2 partial pressure driving force for recarbonation of the acceptor. The CO_2 concentration in the regenerator product gas, as shown by the CO_2 analyzer, was maintained at the programmed value by adjusting the CO_2 and N_2 flows added to the inlet gas.

The CO concentration in the regenerator product gas was followed on the IR analyzer. The air flow rate and fuel char feed rate were adjusted to maintain the desired CO concentration.

The rotameter settings and product gas meter readings were recorded hourly as were the important pressures, pressure drops, temperatures and Powerstat settings.

5. Shutdown Procedure

Just prior to shutdown, the CO_2 flow was removed from the regenerator. The run was ended by simultaneously shutting off the acceptor feeder L-4, the fuel char feeder L-3 and the process air flow, and closing the emergency solenoid valve. The regenerator bed was held at 1800°F until the CO_2 concentration, as shown by the CO_2 analyzer, was essentially zero and then was cooled to 1600°F.

The steam and H_2 flows were removed from the gasifier as soon as possible after the end of the run. Immediately thereafter the gasifier char feed was stopped and the contents of the gasifier were removed via the acceptor withdrawal system.

When the gasifier bed had been removed and the regenerator bed had been cooled to 1600°F, the recycle compressors were shut down and the unit was depressured.

E. Equipment and Procedure for Single-Vessel Operation

Only the reactor, vessel D-1 in Figure 12, was used for single-vessel operation. A simplified flow diagram for this system is shown in Figure 13. The system was used for either preoxidation of coal or for gasification of the preoxidized coal product. A bed of noncaking start-up char was fed to the reactor in both instances.

1. Preoxidation

Coal was fed through the L-6 acceptor outlet line. A thermocouple was installed in the unused overhead char inlet line, and the tar receivers were connected to the system. All of the analyzers were switched to the gasifier.

Preoxidizer outlet gas generally contained some heavy tar and pitch which would rapidly plug the solids filters and the outlet piping. This was alleviated by installing two parallel tar receivers or knock-out pots to collect the heavy material as shown in Figure 13. The upstream piping was heated to prevent condensation of the tar and pitch. The tar receivers were water cooled to ensure condensation of heavy tars; some water was inevitably also condensed. The tar receivers were removed for gasification runs.

To begin preoxidation the bed was brought to temperature with only nitrogen being fed. The coal feed and air were both begun at a reduced rate. The nitrogen flow was diminished in accordance with the increasing air feed. At the same time the electrical heat input was reduced and the temperature was stabilized at the programmed level. Since only a fraction of the final air was being fed, severe temperature excursions were avoided. The final feed rates of coal and air were approached in several steps.

The run was shut down by first turning off the coal feed. The air flow was kept at half rate for 15 minutes while the bed was cooling. This was to prevent possible agglomeration from occurring in the hot bed in the absence of air. Air was then shut off and the bed was drained under nitrogen.

2. Gasification

Preparation of the vessel for gasification was the same as for preoxidation except that the tar receivers were not used. All of the single-vessel runs were used to test the resistance of preoxidized coal to agglomeration and ash slagging, so preoxidized coal was fed from the bottom in an air-containing stream in order to simulate the bottom of the process vessel where the threat of agglomeration or ash slagging would be the greatest.

Gasification conditions were such that the start-up bed completely reacted with any air being fed. Consequently, the programmed air and gas flows were used while the char bed was brought up to temperature. When the start-up char feed hopper ran empty, the programmed feed was switched in and the run began.

Sampling and shutdown were the same as for gasification in the integrated operation.

F. Sample Calculations

1. Preoxidation Runs

The procedure is described below using Run 5P2 as a typical example.

a. Temperature Profile

The preoxidizer bed temperature was continuously measured by three thermocouples at positions corresponding to 10", 16" and 33" above the bottom of the thermowell. Axial bed traverses were taken hourly, and typical traverses are shown in Figure 28. The nomincal bed temperature was taken as the flat zone temperature. In Run 5P2 this was 800°F.

The observed profile is not ideal, but the heat of reaction is of the order of 220,000 Btu/lb-mol, and the fluid bed did not completely smooth out this considerable heat release.

b. Solids Rate - Run 5P2

Feed Rate

In Run 5P2, 52.34 lb of coal were fed over a 6.0 hour period. The coal contained 6.80% moisture, so the average feed rate was:

Dry coal	8.13	lb/hr
Moisture	.592	lb/hr
Total	8.72	lb/hr

The feed rates of the various components of the coal were determined using the ultimate analysis of the coal (dry basis):

Component	Ultimate Analysis Wt % Dry Basis	Input Rate lb/hr
Carbon	60.44	4.91
Hydrogen	4.19	.341
Nitrogen	1.17	.095
Sulfur	4.38	.356
Ash	20.93	1.70
Oxygen (by diff.)	8.88	.722
Volatile Matter	37.21	

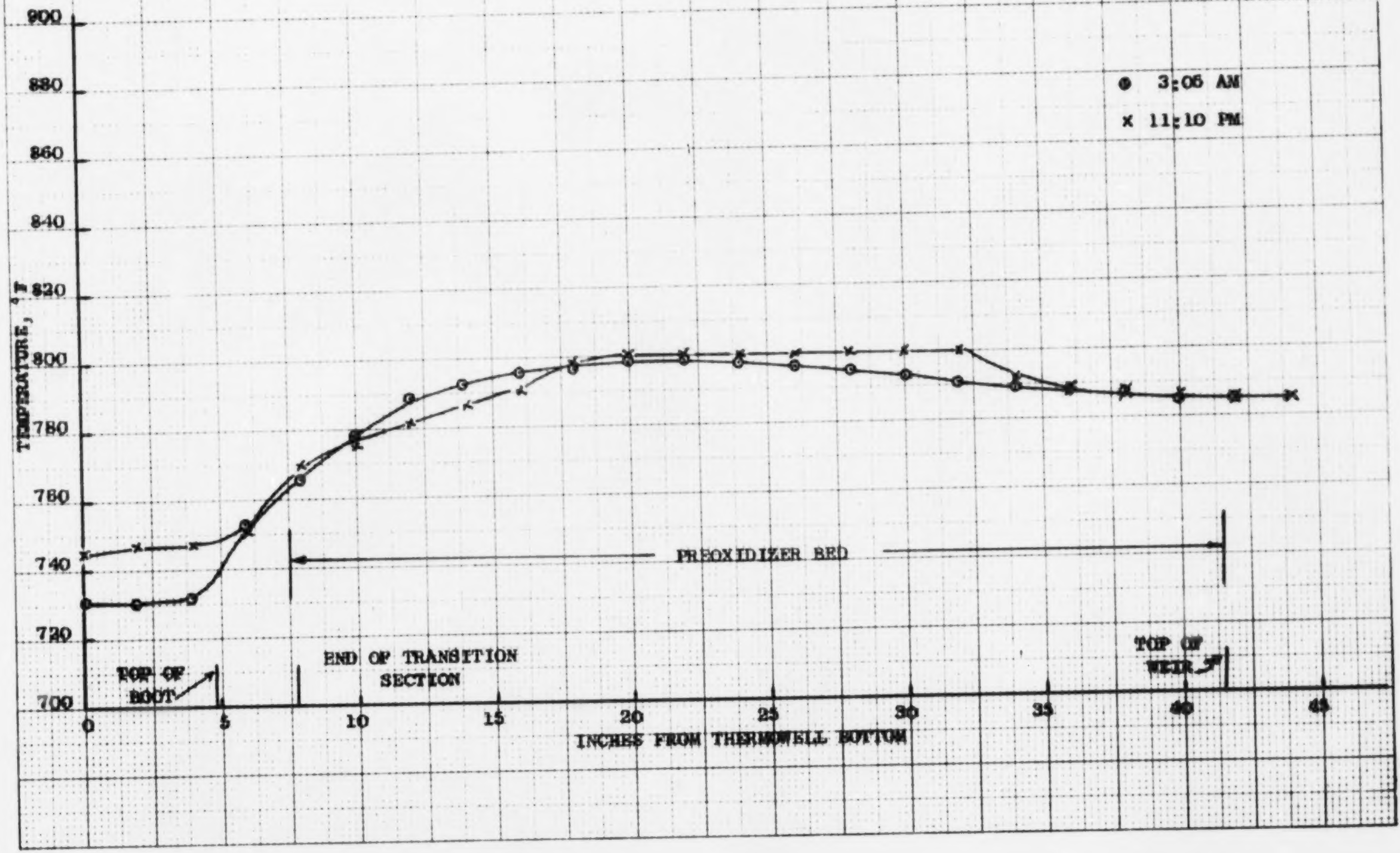
Preoxidized Coal Product

In a five-hour period 34.07 lb of product preoxidized coal were collected. The hourly rate was 6.81 lb/hr and the product was dry when removed and weighed. The feed rates of the various components were determined using the ultimate analysis of the product (dry basis):

Component	Ultimate Analysis Wt % Dry Basis	Exit Rate lb/hr
Carbon	67.03	4.56
Hydrogen	3.29	.224
Nitrogen	1.44	.098
Sulfur	3.72	.253
Ash	16.55	1.13
Oxygen (by diff.)	7.98	.543
Volatile Matter	21.55	

FIGURE 28

PREOXIDIZER TEMPERATURE PROFILE, RUN 5P2



c. Overhead Fines

Overhead fines were collected in two places; the tar receiver, where it stuck to the trapped tarry material in the stainless steel mesh, and in the solids filters. Very little tar, on a weight basis, was actually observed in the tar receiver, and the non-aqueous material was counted as all overhead fines. In Run 5P2, 0.267 lb of solids were removed from the tar receiver and 0.055 lb from the solids filter at the end of the run. This was equivalent to 0.249 lb during the 6 hour balance period or 0.041 lb/hr. The overhead fines were assumed to have the same ultimate analysis as the product coal.

d. Condensate and Tar

Both tar and condensate were collected from each cooler. Over a 6.47 hour balance period the following results were obtained:

Tar + Water	9.51 lb	
Tar	1.00 lb	= 0.155 lb/hr
Water	8.51 lb	= 1.315 lb/hr

Of the 1.315 lb of total condensate, 0.593 were fed in with the coal as moisture and 0.722 were the products of oxidation.

The components exiting with the tar were:

Component	Ultimate Analysis Wt %	Exit Rate lb/hr
Carbon	78.42	0.121
Hydrogen	9.25	0.0143
Nitrogen	0.25	<0.001
Sulfur	5.11	0.0079
Oxygen (by diff.)	6.97	0.0108

e. Gas Flows and Composition - Run 5P2

The inlet flows consisted of air and nitrogen as follows:

	SCFH	Lb/hr
N_2 in air	42.66	3.086
N_2 in lift gas	99.00	7.162
N_2 purges to bed	5.00	0.362
Total N_2 in	146.66	10.61
O_2 in air	11.34	.9376

In addition, there was a recycle stream of 380 SCFH, mostly nitrogen. There were several purges above the bed, and these were determined by difference.

Exit Flow

The exit flow, which was measured with a dry gas meter and converted to standard conditions, was 164.4 SCFH. The gas analysis as obtained with chromatographic and infrared techniques was as follows:

Component	Mol %	lb/hr	C, lb/hr	O, lb/hr
CO_2	3.36	.6267	.1710	.4557
CO	.46	.0546	.0234	.0312
CH_4	.38	.0258	.0193	--
C_2H_6	.14	.0178	.0142	--
C_2H_4	.02	.0023	.0020	--
C_3H_8	.06	.0112	.0092	--
C_3H_6	.01	.0017	.0015	--
COS	.05	.0127	.0024	.0034
SO_2	.18	.0489	--	.0244
N_2 (by diff.)	95.34			

The nitrogen purges above the bed were:

$$.9534 \times 164.4 - 146.7 = 156.7 - 146.7 = 10 \text{ SCFH}$$

The total purges were, therefore: = 15 SCFH

f. Bed Conditions - Run 5P2

The total flow just above the bed consists of the recycle flow, the exit gas flow, and the product steam flow less the purges injected above the bed. For run 5P2 this was:

$$380 + 164 + \frac{1.315 \times 387}{18.02} - 9.7 = 563 \text{ SCFH}$$

The general equation for the fluidizing velocity is as follows:

$$V = \frac{Q}{3600} \times \frac{144}{A} \times \frac{1}{\pi} \times \frac{T}{530}$$

or

$$V = 7.55 \times 10^{-5} \frac{QT}{A\,\pi} \qquad (10)$$

where:

V = fluidizing velocity, ft/sec.
Q = gas flow rate, SCFH.
A = cross-sectional area of vessel, in.^{2}
T = bed temperature, °R.
π = system pressure, atm.

The cross-sectional area of the reactor was 12.20 in.^{2}. Thus,

$$V = 7.55 \times 10^{-5} \frac{563 \times 1260}{12.20 \times 15} = 0.29 \text{ ft/sec.}$$

g. Inlet Oxygen Pressure

This is equal to the fraction of oxygen in the inlet gas times the pressure. The recycle flow is included in the inlet gas flow. The inlet oxygen pressure in Run 5P2 was:

$$\frac{11.34 \times 15}{146.7 + 11.34 + 380} = 0.32 \text{ atm.}$$

h. Extent of Preoxidation

The extent of preoxidation is expressed as pounds of oxygen reacted per pound of dry coal fed times 100. For Run 5P2,

$$\% \text{ Preoxidation} = 0.938 \times 100/8.13 = 11.5\%.$$

i. Material Balances

Carbon Balance

Carbon in on coal .6044 x 8.13 = 4.91 lb/hr.

Carbon in exit streams:

	Lb/hr	Percent
Preoxidized Coal	4.56	92.87
Overhead Fines	.027	.55
Tar	.122	2.48
CO	.023	.48
CO_2	.171	3.48
COS	.002	.04
Hydrocarbons (C_1-C_3)	.046	.94
		100.84

Oxygen Balance

The oxygen balance showed more oxygen in the products than in the entering air. The difference was assumed to come from the coal.

The exit gas was assumed to be saturated at 50°F and 15 atm. Then, in the exit gas:

$$\text{SCFH } H_2O = \frac{.178}{15 \times 14.7} \times 164.4 = 0.133, \text{ or } 0.006 \text{ lb/hr.}$$

	Lb/hr	Percent
To Water		
Condensate	.641	54.7
Moisture in gas	.005	0.4
To CO_2	.456	38.9
To CO	.031	2.7
To SO_2	.024	2.1
To COS	.003	0.3
In Tar	.011	0.9
Total	1.172	100.0

The sources of the oxygen were:

From inlet air	.938 lb/hr
From coal (by diff.)	.234 lb/hr

Coal Balance

	Lb/hr	Percent
Coal In	8.13	--
Preoxidized Coal Out	6.81	83.76
Overhead Fines	.041	.50
Tar	.155	1.91
Carbon in CO + CO_2	.194	2.39
Hydrogen in H_2O	.082	1.00
Hydrocarbons (C_1-C_3)	.059	.72
Sulfur in SO_2 and COS	.031	.38
Coal Oxygen to Products	.234	2.88
Accumulation in Bed (by diff.)	.524	6.46
Total	8.13	100.00

In Run 5P2, the carbon balance was 100%, but the coal balance was only 94%. This was due to the accumulation of dense rocky material in the bed. In Run 6P, the pressure drop across the fluidized bed increased continuously. Microscopic examination and chemical analysis of the final bed material drained from the reactor showed that ash particles had accumulated in the bed during the run. Ash and pyritic sulfur balances on the product coal and bed material showed that 10.55% of the dry feed coal had accumulated in the bed, compared with 10.29% as obtained by difference in the overall weight balance. A check of the chart records for Run 5P2 also showed an increase in pressure drop across the bed, but to a lesser extent than in Run 6P. The accumulation in Run 5P2 was obtained by difference in the overall weight balance, in view of the good agreement between the measured and difference values shown in Run 6P.

2. Gasification Runs

Run 5G3 is used as an example to illustrate the procedure for gasification runs.

a. Temperature Profile

During the run, the temperature was continuously measured by three thermocouples placed at 10", 16" and 33" in the thermowell. Traverses were taken hourly, and typical traverses are shown in Figure 29. The nominal run temperature was taken at the flat zone. For Run 5G3 this corresponded to 1700°F after correction for thermocouple error.

b. Solids Rates

Preoxidized Coal Feed

The feedstock was Run 5P2 composite product, whose analysis was given earlier. During the final balance period, 13 lb were fed in 3.13 hours, or 4.15 lb/hr.

Product Char

The last timed product collection yielded 1.51 lb/hr of product char. This was analyzed to contain 65.64% carbon, 0.46% hydrogen and no oxygen.

c. Gas Rates

Inlet Gas Flows

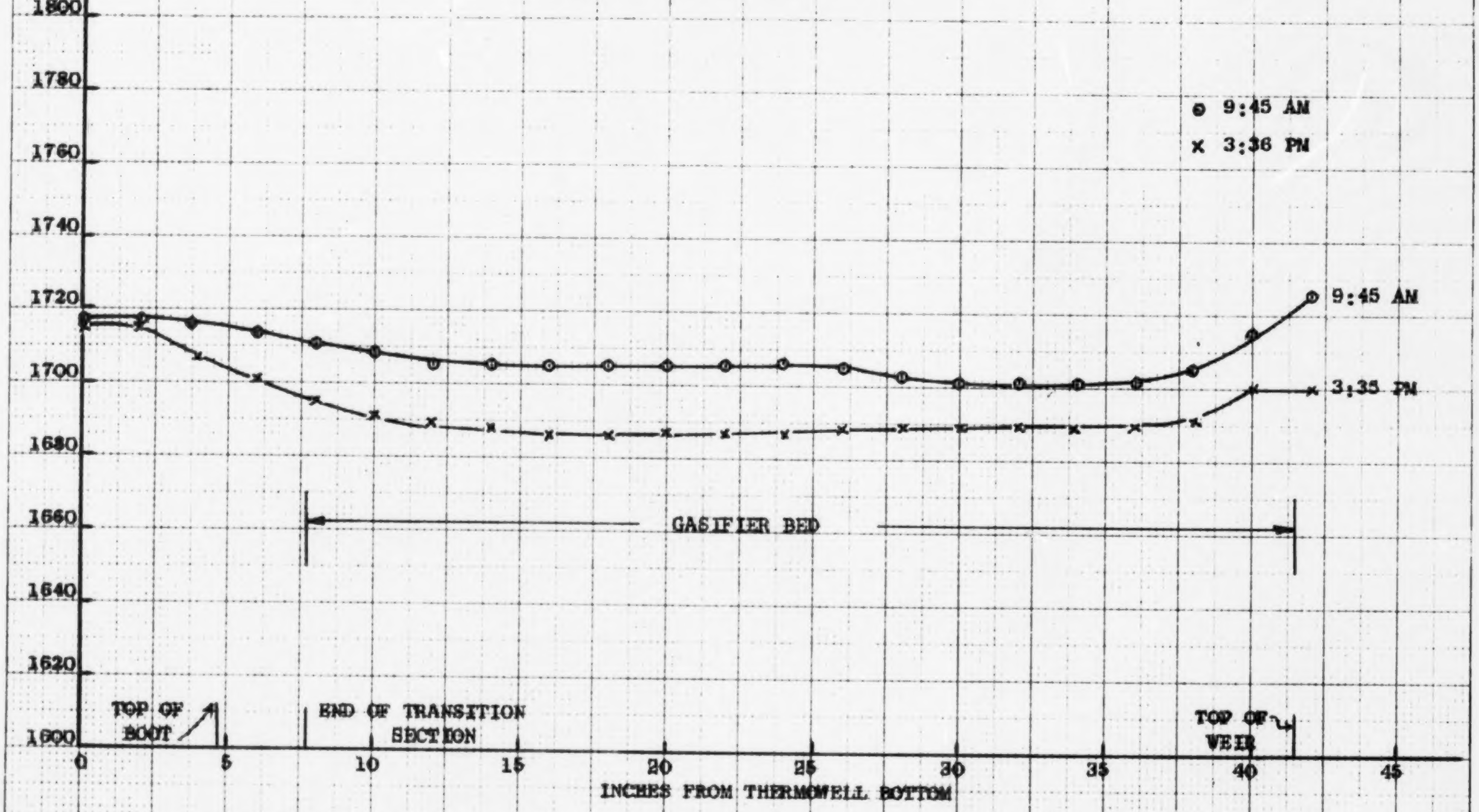
No recycle flows were used. The inlet flows were as follows:

	SCFH
Fluidizing	
CO_2	38
N_2	64
Lift Gas	
Air	105
N_2	34
N_2 purges to bed	5
N_2 purges above bed	6

The fluidizing gas bubbled through a steam generator whose temperature was held at 325°F. At equilibrium, the fluidizing gas would have contained 76 SCFH of steam. Conditions matched those at the bottom of the bed as specified in the process design.

FIGURE 29

GASIFIER TEMPERATURE PROFILE, RUN 5G3



Outlet Gas Flows

The outlet gas rate was measured to be 334 SCFH. Concentrations of CH_4, CO and CO_2 were measured by a gas chromatograph. No analysis was available for the hydrogen concentration. The condensate rate was measured as 2.70 lb/hr. Extensive experience has shown that shift equilibrium is always achieved in the exit gas. The hydrogen concentration was determined from the water gas shift equilibrium, then the inlet steam rate was calculated by forced hydrogen balance. Exit nitrogen was determined by difference. The chromatograph gave the following results for the exit gas:

Component	Mol %
CH_4	.86
CO	13.81
CO_2	14.97

d. Hydrogen Balance and Concentration

In the dry outlet gas, we have:

Component	Mol %	SCFH	SCFH C	SCFH H_2
CH_4	.86	2.88	2.88	5.76
CO	13.81	46.10	46.10	--
CO_2	14.97	49.95	49.95	--
			98.93	

Outlet H_2O = 2.70 lb/hr = 57.95 SCFH.

The total flow just above the bed = dry exit flow + steam - purges above the bed. Thus, the total flow above the bed was:

$$333.7 + 57.95 - 6.0 = 385.6 \text{ SCFH}.$$

On a wet basis just above the bed we get:

Component	Mol %
CH_4	.75
CO	11.96
CO_2	12.95
H_2O	15.03

At 1700°F, the equilibrium constant for the water gas shift reaction is .6973. Then,

$$.6973 = \frac{(H_2)(CO_2)}{(H_2O)(CO)}$$

$$(H_2) = \frac{.6973 \times 15.03 \times 11.96}{12.95} = 9.67\% \text{ or } 37.29 \text{ SCFH}.$$

The total hydrogen leaving was:

		SCFH
as H_2		37.29
in CH_4		5.76
in H_2O		57.95
in Char	$\frac{.0046 \times 1.51 \times 387}{2.016}$	1.33
Total H_2 leaving		102.33

The hydrogen entering with the feed coal is:

$$\frac{.0329 \times 4.15 \times 387}{2.016} = 26.21 \text{ SCFH}$$

The inlet steam rate is therefore 76.12 SCFH

The exit compositions are summarized below:

Component	Mol % (dry)	Mol % (wet)
H_2	11.17	9.67
CH_4	.86	.75
CO	13.81	11.96
CO_2	14.97	12.95
N_2 (by diff.)	57.54	49.64
H_2O	--	15.03

e. Outlet Fluidizing Velocity

The gasifier was the same vessel as the preoxidizer, and as discussed in Section IV-E-1-e, the outlet fluidizing velocity was:

$$V = \frac{7.55 \times 10^{-5} \times 385.6 \times 2160}{12.20 \times 15} = .34 \text{ ft/sec}$$

f. Gasification Rate

Carbon fed = 4.15 x .6703 = 2.78 lb/hr.

$$\text{Carbon gasified} = \frac{(98.93 - 38.00)\ 12}{387} = 1.89 \text{ lb/hr.}$$

The weight of the bed removed from the gasifier at the end of the run was 4.56 lb. The gasification rate is defined as 10^4 x lb carbon gasified/min/lb carbon in bed.

$$\text{Gasification rate} = \frac{1.89 \times 10^4}{4.56 \times .6564 \times 60} = 105$$

Percent carbon burnoff = 100 x 1.89/2.78 = 68.

V. RESULTS AND DISCUSSION

A. Tabular Chronological History of Runs

The purpose, general run conditions, and major conclusions drawn for all the runs are given in the tables listed below. All twenty runs were made at a pressure of 15 atmospheres.

Table No.	
28	Preoxidation Runs
29	Gasification Runs
30	Integrated Gasifier-Regenerator Runs

TABLE 28

Summary of Preoxidizer Operations

Run Number	1P	2P1	2P2
Unit Revisions	None	None	None
Date of Run	5/7/71	5/20/71	5/25-5/26/71
Run Duration, hours	11	2.0	19.0
Purpose of Run	High Pressure Preoxidation of Ireland Mine coal	Preoxidize at 600°F to avoid caking during preoxidation	Preoxidize at 750°F
Temperature, °F	700	600	750
Feedstock	←Ireland Mine Coal→		
Shutdowns	Voluntary	Rapid increase in temperature to 800°F	Voluntary
Results and Conclusions	1. Some caking occurs during preoxidation at 700°F. 2. Product probably not operable in gasifier. 3. No oxygen breakthrough. 4. 19.5% preoxidation.	1. Distortion of temperature profile indicated agglomeration probably occurred. 2. 19% preoxidation.	1. No agglomeration or caking observed. 2. Startup procedure changed to get oxidation of start up char to occur before adding coal. 3. 18.6% preoxidation.

TABLE 28
(Continued)

Summary of Preoxidizer Operations

Run Number	3P	3P1	4P	4P1
Unit Revisions	Added pitch and heavy tar receiver for use in preoxidation. Replaced L-2 feeder system with ball valve.	None	None	None
Date of Run	6/3/71	6/3-6/4/71	6/4/71	6/4/71
Run Duration, hours	1.0	2.0	3.75	
Purpose of Run	Preoxidize at 800°F to improve operability of product in the gasifier.	Repeat 3P approaching 800°F more slowly.	Find preoxidation conditions which would make Ireland Mine coal operable in gasifier.	Effect of 800°F preoxidation.
Temperature, °F	← 800 →		750	800
Feedstock	← Ireland Mine Coal →		← Run 2P2 Composite Product →	
Shutdowns	Distortion of temperature profile indicated abnormal mixing pattern	Bed caked	Change to 4P1 conditions	Very low density material would not flow reliably through overflow weir. Reactor filled and plugged gas exit line.
Results & Conclusions	1. Large agglomerates found in reactor. 2. 20% preoxidation.	1. Bed filled with agglomerates. 800°F is not an operable preoxidation temperature. 2. 20% preoxidation.	1. Product preoxidized to 28% total.	1. No agglomerates formed, but product had a very low density.

TABLE 28
(Continued)

Summary of Preoxidizer Operations

Run Number	5P1	5P2	6P
Unit Revisions	None	None	Installed bellows on gasifier tailpieces.
Date of Run	8/2/71	8/3-8/4/71	8/11/71
Run Duration, hours	9.25	6.0	5.5
Purpose of Run	Preoxidize Hillsboro coal at 750°F.	Preoxidize Hillsboro coal at 800°F.	Preoxidize Hillsboro coal at process conditions: 8.5% preoxidation, 810°F.
Temperature, °F	750	800	810
Feedstock	Hillsboro Coal	Hillsboro Coal	Hillsboro Coal
Shut downs	Plug in coal feed line.	Could not switch tar receivers due to plug in receiver inlet line.	Voluntary.
Results and Conclusions	1. No operability problems encountered. 2. 11.9% preoxidation.	1. No operability problems encountered. 2. 11.5% preoxidation.	1. No operability problems encountered. 2. 8.7% preoxidation.

TABLE 29

Summary of Gasifier Operations

All runs at 1700°F

Run Number	OG2	OG3	OG4	OG5
Unit Revisions	None	None	Char and air inlet raised into bed	None
Date of Run	5/13/71	5/13/71	5/13/71	5/13/71
Run Duration, hours	9.5	2.0	9.0	2.0
Purpose of Run	Gasification at 0.43 atm., O_2 inlet	Gasification at 1.04 atm., O_2 inlet	Gasification at 1.0 atm., O_2 inlet	Gasification at 2.4 atm., O_2 inlet
Feedstock	← Partially Gasified Disco Char →			
Shutdowns	Switch to OG3 Conditions	Voluntary	Switch to OG5 Conditions	Voluntary
Results and Conclusions	← No evidence of ash slagging →		← No evidence of ash slagging →	

TABLE 29
(Continued)

Summary of Gasifier Operations

All runs at 1700°F

Run Number	1G3	2G3	4G3	5G3	6G3
Unit Revisions	Char and Air Inlet to Gasifier Bottom	None	None	None	None
Date of Run	5/19/71	5/27/71	6/7/71	8/5/71	8/13/71
Run Duration, hours	0.22	0.75	2.5	8.0	7.6
Purpose of Run	Test 1P Product for operability at OG3 conditions.	Test 2P2 product for operability at OG3 conditions.	Test 4P product at OG3 conditions.	Test operability of 5P2 product in gasifier.	Test 6P1 product in gasifier.
Feedstock	Run 1P composite product	Run 2P2 composite product	4P product	5P2 product	6P product
Shutdowns	Bed caked	Voluntary	Leak in char exit line	Voluntary	Voluntary
Results and conclusions	1. Run 1P product is not operable at OG3 conditions.	1. No caking, but significant agglomeration occurred.	1. No caking 2. No ash slagging	1. No agglomerates 2. No ash slagging	

TABLE 30

Results of Integrated Gasifier-Regenerator Operation

Run Number	D1
Unit Revisions	Installed acceptor sample line between regenerator and gasifier
Date of Run	8/23/71 and 8/30/71
Purpose of Run	Demonstrate integrated operation of regenerator-gasifier at process conditions
Temperature:	
Gasifier	1700° F
Regenerator	1860° F
Feedstock:	
Gasifier	Disco Char Tymochtee 9 dolomite
Regenerator	Pregasified Disco Char
Shutdowns:	Cycle 5 - Replaced corroded sample valve, inspected vessels. Delayed resumption to attend OAP contractors meeting. Cycle 13 - Regenerator thermowell failed due to localized sulfur attack.
Results and Conclusions	1. No ash slagging in either vessel. 2. No agglomeration or transient liquid formation in regenerator. 3. Reaction of CO_2 with carbon in the regenerator may be slower than predicted.

B. Preoxidation Runs

1. Introduction and Tabulated Data

A major aspect of the adaptation of the CO_2 Acceptor Process for making low-sulfur boiler fuel gas is the use of Eastern steam coals. These coals become fluid when heated through the temperature range of roughly 700-850°F, and therefore require pretreatment by preoxidation to prevent caking at gasification conditions.

Numerical data for run conditions and results of all operable preoxidation runs are given in Table 31. Analytical data on the properties of the coal feed and products are given in Table 32. Material balance data and distribution of oxygen in the products are given in Tables 33 and 34, respectively.

Other tabulated data specific to the topic being discussed are presented in the appropriate succeeding sections.

Typical relationships among the intensive variables, severity of preoxidation and inlet oxygen partial pressure, and the extensive variables, coal feed rate and nominal residence time are shown in Figure 30. The data shown were obtained from straightforward material balance calculations, assuming all the inlet oxygen reacts. Specific conditions used in the calculations are shown in Figure 30. The reactor used during the runs has an inside diameter of 4.0 inches, with a bed volume of 0.25 ft^3.

TABLE 21

Preoxidation Conditions and Results

System Pressure: 15 atmospheres (206 psig)
Size Consist : 24 x 100 Tyler Mesh

Run Number	1P	2P2	4P	5P1	5P2	6P
Feed Coal	Ireland Mine		2P2 Product	Illinois No. 6		
Temperature, °F	700	750	750	750	800	810
Inlet O_2 Partial Pressure, atm	.49	.40	.32	0.31	0.32	0.25
Fluidizing Velocity (top of bed), ft/sec	.34	.34	.29	0.29	0.29	0.29
Moisture in Coal, Wt %, as fed	1.58	1.58	1.03	6.80	6.80	6.38
Input						
Coal Feed Rate, lb/hr (dry basis)	6.5	7.46	9.34	7.89	8.13	8.71
Lift Gas, SCFH						
Air	73	80	54	54.0	54.0	43.5
N_2	84	70	96	97	99	110
Fluidizing Gas, SCFH						
Recycle	310	390	380	396	380	382
N_2	0	92	0	0	0	0
Purges (N_2), SCFH to bed	5	5	5	5	5	5
Purges (N_2), SCFH above bed	15.5	15.5	15.5	10	9	6
Output						
Exit Gas Rate, SCFH(1) (dry basis)	169	256	159	161	164	161
Exit Gas Composition, Mol % (dry basis)						
O_2	<.02	<.02	<.02	<.01	<.01	<.01
CO_2	3.53	2.56	3.33	3.03	3.36	2.54
CO	1.20	.85	.89	.40	.46	.25
CH_4	--	.22	.21	.25	.38	.45
C_2H_6	--	--	--	.08	.14	
C_2H_4	--	--	--	.01	.02	
C_3H_8	--(3)	--(3)	--(3)	.01	.06	(5)
C_3H_6	--	--	--	.01	.01	
COS	--	--	--	.02	.05	
SO_2	--	--	--	.16	.18	.09
H_2S	--	--	--	<.003	<.003	--
H_2	--	--	--	<.01	<.01	--
N_2 (by difference)	95.3	96.4	95.6	96.03	95.34	96.39
Flow Rate, SCFH, at top of bed	500	669	552	572	563	569
Water, lb/hr (corrected for coal moisture)						
Condensate	.665	.729	.265	.390	.722	.824
Moisture in Exit Gas	.006	.010	.006	.006	.006	.006
Tar, lb/hr	0	.0149	0	.0934	.155	.292
Preoxidized Coal, lb/hr	(2)	6.96	9.14	7.20	6.81	6.76
Percent Preoxidation	19.5	18.6	10.0(4)	11.9	11.5	8.7
Nominal Solids Retention Time, min	(2)	57	49	59	46	39
Duration of Steady-State Period, hr	11	19	3.75	10.0	6.0	5.5
Total Product, Pounds	70	132	34	72	41	37

(1) Standard conditions are 29.92 in Hg and 70°F.
(2) Uncertain because of agglomerate in reactor.
(3) Not measured.
(4) 9.3% based on original raw coal.
(5) Not measured. Assumed to be the same as in Run 5P2.

TABLE 32

Properties of Coal and Products - Preoxidation Runs

Run Number	Pittsburgh Seam Ireland Mine Feed Coal		1P		2P2		4P	
Hydrogen, Wt % (dry basis)	5.00		4.14		3.74		3.26	
Carbon	72.24		70.51		69.99		69.81	
Nitrogen	1.35		1.45		1.44		1.43	
Oxygen (by difference)	5.53		8.75		8.07		7.99	
Sulfur	4.52		4.03		3.93		3.75	
Ash	11.36		11.12		12.83		13.76	
Volatile Matter	40.1		32.7		28.6		24.7	
Pyritic Sulfur	2.53		1.76		1.73		1.57	
Sulfate Sulfur	.07		.08		.10		.11	
Size, Tyler Mesh	Wt %	Density[1]	Wt %	Density	Wt %	Density	Wt %	Density
+24	0	81.0	3.4	42.0	4.0	41.2	2.6)	47.7
24 x 28	11.2		16.3	50.6	16.1	42.4	13.8)	
28 x 35	35.4		31.0	57.5	27.3	51.5	24.2	51.5
35 x 48	34.8		31.1	71.2	31.0	56.5	29.2	58.8
48 x 65	16.9		15.9	73.9	15.2	62.8	16.9	61.9
65 x 100	1.2		1.8	76.8	4.3	73.1	7.8	70.2
-100	0.5		0.5	76.8	2.1	74.0	5.5	70.9
Mean Diameter, in[2]	.0165		.0174		.0169		.0156	
Mean Density, lb/ft^3[3]	81.0		61.6		53.1		56.4	

(1) Measured in mercury at 1 atmosphere.
(2) Arithmetic mean.
(3) Reciprocal mean.

TABLE 32
(Continued)

Properties of Coal and Products - Preoxidation Runs

Run Number	Hillsboro Mine Illinois No. 6 Feed Coal (1)		Hillsboro Mine Illinois No. 6 Feed Coal (2)		5P1		5P2		6P	
Hydrogen, Wt % (dry basis)	4.19		4.18		3.18		3.29		3.44	
Carbon	60.44		60.14		61.72		67.02		70.61	
Nitrogen	1.17		1.22		1.34		1.44		1.36	
Oxygen (by difference)	8.81		7.67		7.81		7.98		5.80	
Sulfur	4.46		4.93		4.07		3.72		3.40	
Ash	20.93		21.86		21.88		16.55		15.39	
Volatile Matter	37.2		36.4		23.4		21.6		22.0	
Pyritic Sulfur	2.20		2.26		1.51		1.11		1.37	
Sulfate Sulfur	.07		.30		.33		.27		.07	
Size, Tyler Mesh	**Wt %**	**Density**	**Wt %**	**Density**	**Wt %**	**Density**	**Wt %**	**Density**	**Wt %**	**Density**
+24	0		0		0.4		1.4	37.9	2.0)	50.0
24 x 28	9.6		15.8		7.9	52.3	14.0	40.1	17.0)	
28 x 35	39.3		41.1		33.3	57.0	35.5	44.1	34.5	50.5
35 x 48	34.2		29.0		34.7	60.5	32.0	58.8	29.4	49.2
48 x 65	15.0		12.7		20.2	77.1	15.0	61.5	14.3	64.3
65 x 100	1.3		1.0		3.0	81.2	1.8	72.9	2.2	58.9
-100	0.6		0.4		0.5		0.4		0.5	
Mean Diameter, inch(3)	.0166		.0176		.0158		.0171		.0175	
Mean Density, lb/ft^3(4)	80.0		80.0		61.7		49.8		51.8	

(1) Batch No. 1, used in Runs 5P1 and 5P2.
(2) Batch No. 2, used in Run 6P.
(3) Arithmetic mean.
(4) Measured in mercury at 1 atmosphere. Reciprocal mean.

TABLE 33

Material Balances for Preoxidation Runs

Basis: 100 lb Dry Coal Fed

Run Number	2P2	4P	5P1	5P2	6P
Preoxidized Coal	93.30	97.85	91.25	83.76	77.60
Overhead Fines	.58	1.21	.70	.50	.91
Tar	.20	0	1.18	1.91	3.35
Carbon in $CO + CO_2$	3.65	2.23	2.16	2.39	1.60
Hydrogen in H_2O	1.10	.33	.84	1.00	1.06
Hydrccarbons (C_1-C_3)	.34(1)	.15(1)	.40	.72	.72
Sulfur in SO_2 and COS	.75(2)	.26(2)	.30	.38	.22
Coal Oxygen to Products	0	0	.61	2.88	4.25
Accumulation in Bed, Ash and S as FeS_2	0	0	2.56(3)	6.46(3)	10.29(3,4)
Total	99.92	102.03	100.00	100.00	100.00

(1) Only CH_4 was determined.
(2) By sulfur balance on coal feed and preoxidized coal product.
(3) By difference.
(4) Ash and sulfur balances on product and bed materials gave 10.55 lb accumulation.

TABLE 34

Distribution of Oxygen in Products of Preoxidation

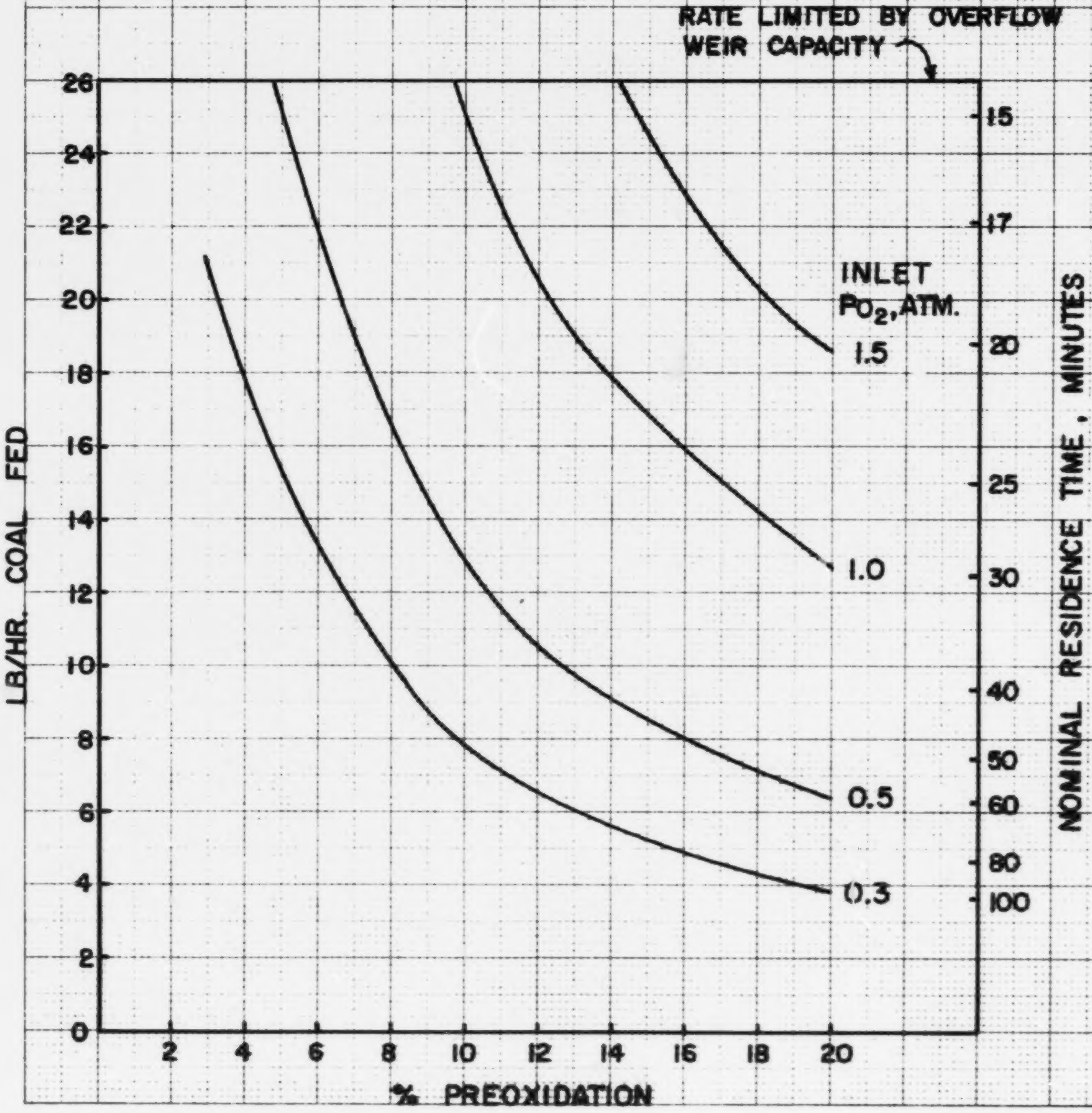
Run Number	2P2		4P		5P1		5P2		6P	
	Lb/hr	Percent	Lb/hr	Percent	Lb/hr	Percent	Lb/hr	Percent	Lb/hr	Percent
Output										
To Water										
Condensate	.648	46.6	.235	25.1	.524	53.2	.641	54.7	.732	65.0
Moisture in Gas	.009	.6	.006	.6	.005	0.5	.005	0.4	.005	0.4
To CO_2	.546	39.3	.437	46.7	.402	40.8	.456	38.9	.338	30.1
To CO	.090	6.5	.058	6.2	.027	2.7	.031	2.7	.017	1.5
To SO_2	.056	4.0	.024	2.5	.021	2.1	.024	2.1	.012	1.1
To COS	[(1)]	-	[(1)]	-	.001	0.1	.003	0.3	.003	0.3
In Tar		-		-	.006	0.6	.011	0.9	.018	1.6
On Coal (by diff)	.040	3.0	.178	18.9	0	0	0	0	0	0
Total	1.389	100.0	.937	100.0	.986	100.0	1.172	100.0	1.125	100.0
Input										
From Coal (by diff.)	0		0		.048		.234		.370	
Lb/hr O_2 from Air	1.389		.937		.938		.938		.755	

(1) Not determined.

FIGURE 30

RELATIONSHIPS AMONG PREOXIDATION VARIABLES

SYSTEM PRESSURE	15 ATM.
TEMPERATURE	700° F
FLUIDIZING VELOCITY	.24 FT/SEC.
PARTICLE DENSITY	65 LB/FT.^{3}



2. Pittsburgh Seam Coal (Ireland Mine)

For the initial experimental program with Ireland Mine coal, a severe level of preoxidation* (20%) was chosen deliberately with the intent of suppressing completely any tendency toward caking at gasification conditions. A long nominal retention time (~ 60 minutes) was chosen to insure adequate exposure to O_2 for those particles at the low end of the retention time distribution which is characteristic of a continuously fed fluidized bed. With these constraints, the inlet O_2 partial pressure fell in the range of 0.4-0.5 atm. This level of O_2 partial pressure is representative of that in an upper portion of the envisioned process preoxidizer in which the inlet O_2 partial pressure is 3.15 atm at 15 atm system pressure.

The first three runs were made at conditions shown below:

Run No.	1P	2P1	2P2
Pressure		15 atm (206 psig)	
Temperature, °F	700	600	750
Coal Feed Rate, lb/hr	6.5	7.1	7.46
Fluidizing Velocity, ft/sec	.24	.30	.34
Inlet O_2 Partial Pressure, atm	0.5	0.4	0.4
Percent Preoxidation	←	~ 20	→

Run 1P

Run data and properties of the preoxidized coal product are listed in Tables 31 and 32. The run proceeded without apparent difficulty, except for a disturbingly large temperature gradient across the fluidized bed, until the desired amount of product was obtained. Coal, air and recycle fluidizing gas were fed to the bottom of the preoxidizer at substantially room temperature. During the run, the temperature at the bottom typically was 580°F and the profile showed an increase to 700°F at about 25" above the bottom with constant temperature to the top of the bed at 40".

On disassembly of the reactor, it was found to be partially choked with large chunks of agglomerated coal. Presence of the agglomerates distorted the normal mixing pattern of the fluidized bed and thus accounted for the large temperature gradient. Further, the volume occupied by agglomerated coal decreased the nominal residence time of the coal particles by an unknown amount. However, all the inlet O_2 had been consumed, i.e., the desired 20% preoxidation level had been achieved.

A laboratory test (see page 133) showed that the product had retained much of its caking tendency and that it almost certainly would have caked at gasification conditions. At the time, this situation was ascribed to the low nominal retention time which allowed some of the coal to pass through the preoxidizer without reaching the average level of 20% preoxidation. This material was set aside temporarily.

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* Severity of preoxidation is expressed as percent preoxidation:
100 (lb O_2 consumed/lb dry coal).

Run 2P1

Since coke had formed at 700°F in Run 1P, the next preoxidation run was made at 600°F at the same 20% preoxidation level, in a continued effort to produce an operable feedstock for the gasifier which had had a known history of retention time in the preoxidizer.

Coals with fluidity characteristics similar to Ireland Mine coal historically have been fluidized at atmospheric pressure and 600°F in various Consol pilot plant operations, with and without the presence of O_2, without occurrence of caking or agglomeration.

In Run 2P1, a large temperature gradient developed soon after coal feeding was started. After two hours of feeding, the temperature at 16" above the bottom increased suddenly to 800°F at which time the run was ended. Inspection of the vessel after disassembly showed no evidence of agglomeration. Yet, the temperature gradient and sudden rise, clearly were indicative of disruption of the fluidized bed mixing pattern as if agglomeration had occurred.

To help explain the behavior the following laboratory tests were made. Samples of of raw Ireland Mine coal were heated in a thermobalance purged with helium at atmospheric pressure. The coal reached the desired temperature in about two minutes and was held for an additional 15 minutes. Results are shown below:

	Observations		
Temperature, °F	Particle Rounding	Agglomeration	Coking
650	No	No	No
700	Yes	No	No
750	Yes	Yes	No
800	--	--	Yes

The above run data indicate that no agglomeration should have occurred in Run 2P1, even at 650°F.

Samples of the raw coal were then heated in a small autoclave purged with helium at 15 atm (206 psig) with the following results:

Temperature, °F	Particle Rounding	Agglomeration
600	No	No
630	Yes	Yes

Thus, at 15 atm pressure the coal agglomerates at a temperature below that at which no trace of agglomeration occurs at 1 atm.

In Run 2P1 the particle surface temperature probably reached 630°F while undergoing preoxidation, with consequent formation of agglomerates. This behavior at 600°F is attributed to small amounts of waxy and resinous material which are present in nearly all bituminous coals. This material melts at relatively low temperatures and evaporates at atmospheric pressure before agglomerates can form. At elevated pressure, evaporation is hindered sufficiently so that the particles can stick to each other. The agglomerates so formed must have been very fragile and were broken up during manipulation of the reactor on disassembly after Run 2P1.

Run 2P2

The results of Run 2P1 and the laboratory tests showed the probable cause of agglomeration in Run 1P. During startup, the initial bed was an inert char which had been partially gasified at 1775°F in a previous experimental program. At the 20% preoxidation level, heat must be dissipated through the preoxidizer vessel wall. The start-up procedure was to supply heat electrically to the preoxidizer while feeding the inert char and to bring the wall temperature to that estimated as sufficient to dissipate the strongly exothermic heat of reaction when the feed was switched to coal. Generally, a bed-to-wall ΔT at 60°F was adequate while preoxidizing coal with no electrical heat input to the reactor. At the end of the start-up period, the inert char bed temperature was about 500°F. The char was unreactive toward the inlet O_2 at this temperature. In Run 1P the temperature of the upper part of the bed reached 700°F in about one hour after the coal feeding was started, but the bottom of the bed stayed at about 600°F. It is possible that agglomerates formed while the bed was passing through the 600-630°F range during the heat-up from 500°F. The resulting poor mixing caused a permanent temperature gradient to be established in which several inches near the bottom of the bed were always in the 600-630°F range. Therefore, it is also probable that agglomerates continued to form throughout the run.

In Run 2P2, made at 750°F, the start-up procedure was revised so that no coal ever would be exposed to temperatures in the range of 600-630°F. The inert char bed was brought to about 725°F by means of electrical heat. At this temperature the char was reactive toward the inlet O_2 and relatively little electrical heat was needed. The electrical heat was turned off and, in quick succession, the air flow rate was cut to one-half the programmed flow and coal feeding was started, also at one-half the programmed flow. The flow rates were increased stepwise to the full programmed rates while keeping the air/coal ratio constant. During this period, the bed temperature rose to 750°F and subsequently was held at that level by varying the electrical heat input. The maximum temperature gradient was about 30°F and the temperature profile was essentially flat over the range of 16 to 40" above the bottom. The run was ended voluntarily when the desired amount of product was obtained. Inspection of the vessel after disassembly showed no agglomeration whatever. Run data, some properties of the preoxidized coal product, the distribution of reacted O_2, and a material balance are presented in Tables 31 through 34.

Both Runs 1P and 2P2 products were fed to the gasifier at process conditions. The Run 1P coal melted and coal feeding had to be stopped after 13 minutes. The Run 2P2 material produced numerous small agglomerates. Examination of the gasifier product showed that the coal was sticking to itself and to the Disco char particles in the start-up bed.

It was apparent that at the pressure used in the process, even 20% preoxidation was ineffective in preventing caking in the gasifier. Pretreatment by preoxidation involves dehydrogenation and devolatilization. Elevated system pressure inhibits these processes, while increased preoxidation temperature enhances these same processes. Since a higher preoxidation temperature can compensate for the suppressing effect of elevated pressure on dehydrogenation and devolatilization, a preoxidation run at 800°F was scheduled.

Run 3P

This run was made at the same conditions as for Run 2P2, except that the initial temperature of the start-up bed of Disco char was increased to 775°F by decreasing the heat losses via the electrical heaters. As before, coal feeding was begun at one-half the programmed flows of coal and air. A stepwise increase to two-thirds of the programmed flows brought the temperature to 800°F. Over a period of about one hour the temperature of the middle portion of the bed increased to 850°F, a positive indication that the normal mixing pattern of the fluidized bed had been disrupted.

On inspection of the vessel after disassembly, it was found to be partially choked with large chunks of agglomerates consisting of coal particles which had adhered to themselves as well as to the Disco char particles.

Run 3P1

In this run the initial temperature of the start-up bed was 725°F, as in Run 2P2. During stepwise increases of the coal and air flows, the bed temperature was increased to 800°F by increasing the amount of electrical heat supplied. On reaching full programmed flows after about two hours of feeding coal, the pressure drop across the fluidized bed increased suddenly, an indication of a massive blockage.

After disassembly of the vessel, it was found to be nearly full of agglomerates similar in appearance to that in Run 3P.

Run 4P

To establish a benchmark as to the severity of preoxidation for 24 x 100 mesh Ireland Mine coal which assuredly would provide an operable feedstock for the gasifier, the Run 2P2 product was subjected to a second stage of preoxidation at 750°F in which an additional 10% preoxidation occurred.

The run proceeded without incident until the desired amount of product was obtained. Run data, some properties of the preoxidized coal product, the distribution of reacted O_2, and a material balance are presented in Tables 31 through 34.

The smaller mean particle diameter of the Run 4P product, compared with the feedstock, was caused by decrepitation of some of the rounded, swollen particles on passage through the rotary feeder. The high particle density of the Run 4P product is the result of loss of volatile matter.

As shown in Table 34, during a second stage of preoxidation, much less water forms, in accordance with previous work, and a considerable amount of the reacted O_2 is retained in the coal.

Run 4P1

In order to obtain a gasifier feedstock which was more highly devolatilized, at the end of the 750°F operating period of Run 4P, the bed temperature was raised to 800°F via the electrical heaters while holding all input flows constant.

Increased particle swelling was apparent immediately. After about one hour at 800°F, the very low density material would not flow reliably through the overflow weir with the result that the reactor filled completely with coal. The exit gas line then became plugged with solids. Inspection of the vessel after disassembly showed that no caking or agglomeration had occurred.

However, such low density material (particle density <30 lb/ft^3) would be impractical as a gasifier feedstock from the standpoint of maintaining an adequate carbon inventory.

Run 4P product was then fed to the gasifier at the standard process conditions with no evidence of either caking or ash slagging.

3. Conclusions - Pittsburgh Seam Coal

Work with Ireland Mine coal showed that elevated system pressure has an adverse effect on operability in the preoxidizer as well as on operability of the preoxidized coal in the gasifier. Attempts to preoxidize raw coal at 800°F at the severe level of 20% preoxidation were unsuccessful. During a second-stage preoxidation at 800°F, an intolerable amount of particle swelling occurred, although there was no caking or agglomeration.

After two stages of preoxidation at 750°F to a total severity level of 28%, the Ireland Mine coal was operable at gasifier conditions with no caking or ash slagging. At this level of preoxidation a very large amount of heat would have to be removed from the process preoxidizer, with the attendant economic penalty due to lowered thermal efficiency. For relatively coarse, 24 x 100 mesh, Ireland Mine coal the work to date has not yielded an economically satisfactory preoxidation procedure. Further research is required to find techniques which will decrease the severity of preoxidation to lower and more economic values.

Work is planned which will investigate the possibilities involved in using a two temperature staged preoxidation. The magnitude of the economic penalty involved in operating at high levels of preoxidation will also be investigated.

Additional areas for study are the use of a small size consist of the feed coal and consideration of a continuously temperature staged entrained solids reactor which would adiabatically reach high levels of preoxidation. Still a final possibility is the use of other pretreatment methods which do not involve preoxidation.

4. Illinois No. 6 Coal (Hillsboro Mine)

The inability to make Ireland Mine coal noncaking using process conditions of preoxidation, led to the investigation of the preoxidation behavior of a less severely caking coal. Three runs were made at the conditions listed below:

Run Number	5P1	5P2	6P
Temperature, °F	750	800	810
Inlet O_2 Partial Pressure, atm	0.31	0.32	0.25
Fluidizing Velocity, ft/sec (top of bed)	0.29	0.29	0.29
Percent Preoxidation	11.9	11.5	8.7

Run data, properties of the preoxidized products, material balances and distribution of oxygen in the products are given in Tables 31 through 34.

Preoxidation Runs 5P1 and 5P2

Since these runs were the first to be made with Illinois No. 6 coal, a preoxidation level higher than the adiabatic level was chosen deliberately (10% versus 8.5%).

Upon completion of the 750°F portion of Run 5P1, the temperature was raised to 800°F via the electrical heaters in order to obtain a more highly devolatilized product. Both runs were completely operable.

The extent of preoxidation in these runs was somewhat higher than the desired 10% because the coal feed rate was lower than the calibrated value, due to a worn out variable speed drive.

Routinely, the product gas was analyzed by an on-line gas chromatograph for carbon oxides, methane and nitrogen. In these runs, separate gas samples were taken for analyses in the laboratory for C_2-C_3 hydrocarbons, H_2, COS, and H_2S. Also, the continuous infrared SO_2 analyzer had been received and was used for the first time.

In contrast with Run 2P2 in which Ireland Mine coal was preoxidized at 750°F, Run 5P1 (also at 750°F) gave a 1.2% yield of a heavy tar rather than a 0.2% yield of light oil. It is not clear whether this difference was caused by the lower extent of preoxidation in Run 5P1 (11.9% versus 18.6%) or by an intrinsic difference in the kinetics of formation and of thermal decomposition of the liquids which form when the Illinois No. 6 and Pittsburgh Seam coals are heated to 750°F. A run with Ireland Mine coal at the same conditions used in Run 5P1 may provide a clue as to why preoxidation of Pittsburgh Seam coal is ineffective at elevated system pressure.

Run 6P - Preoxidation at Adiabatic Conditions

In this run the air input and the defective feeder drive were adjusted to give the preoxidation level corresponding to adiabatic conditions in the process preoxidizer. The temperature was increased slightly to 810°F, compared with Run 5P2 in which the temperature was 800°F. The feeder drive since has been replaced.

In Run 6P, the pressure drop across the fluidized bed increased continuously. Microscopic examination and chemical analysis of the final bed material drained from the reactor showed that ash particles had accumulated in the bed during the run. Ash and pyritic sulfur balances on the product coal and bed material showed that 10.55% of the dry feed coal had accumulated in the bed compared with 10.29% as obtained by difference in the overall weight balance, as shown in Table 33. A check of the chart records for Run 5P2 also showed an increase in pressure drop across the bed, but to a lesser extent than in Run 6P. In Table 33 the accumulations in Runs 5P1 and 5P2 were obtained by differences in the overall weight balances, in view of the good agreement between the measured and difference values shown in Run 6P.

The nominal retention times shown in Table 31 are corrected for the presence of accumulated ash.

All of the runs discussed here were completely operable, with no traces of agglomeration in the preoxidizer. The preoxidized coal products from Runs 5P2 and 6P were fed to the gasifier operated at full process air flow as in Runs 5G3 and 6G3, respectively. Both runs were completely operable, with no traces of agglomeration or of ash slagging.

5. Conclusions - Illinois No. 6 Coal

Current results with Illinois No. 6 coal show that this material is completely operable at 750 and 900°F in the preoxidizer at about the 8-12% level of preoxidation.

The coal was run through both preoxidation and gasification at process conditions with no operability problems whatsoever. It appears that there will be no oxygen breakthrough in the process preoxidizer.

6. Laboratory Screening for Operability in the Gasifier

It is useful to have a screening test which can identify those coals which will not be operable in the gasifier. The test should be rapid and provide a quantitative measure of the relative tendency of the coal to cake. Several tests were used:

a. Gieseler Plastometer

The Gieseler plastometer (ASTM Designation D-2639-67T) often is used to measure the caking-coking tendency of coals. In effect the method measures, at standardized conditions, the coal's viscosity while being heated to about 900°F. The coal sample (5 grams) is compacted around a small stirrer to which a constant torque is applied. The sample is heated from 590°F at 5.4°F/minute and the number of rotations of the stirrer shaft is read out continuously on a dial. The coal's fluidity (reciprocal viscosity) is given by the rpm of the stirrer shaft, expressed as dial divisions per minute (DDPM). The fluidity increases with increasing temperature, goes through a maximum, then decreases to zero as thermal decomposition causes the coal to form a rigid mass of coke.

The plastometer data for raw Ireland Mine coal and several of its preoxidized products are plotted in Figure 31. The relationship between fluidity (DDPM) and physical changes which occur when raw coal is heated usually are stated as shown below. The temperatures at which the phenomena occur for raw Ireland Mine coal are also shown:

Physical Change	Appearance of Coal	DDPM	Temp., °F
Incipient melting	Rounding of sharp edges.	1	710
Agglomeration	Coherent agglomerate, no loss of particle identity.	5	742
Fusion	Coke, complete loss of particle identity.	Max.	803

These phenomena, and their associated temperatures, correspond closely with the observations made when Ireland Mine coal was heated in a thermobalance purged with helium at atmospheric pressure (page 124).

As shown in Figure 31, the Run 1P product, made at 700°F, retained considerable fluidity whereas the Run 2P2 product, made at 750°F, had a lower fluidity. The plastometer information is in accord with the results of the gasification runs in which the Run 1P product melted with complete loss of particle identity and the Run 2P2 product formed agglomerates with no loss of particle identity.

Also shown in Figure 31 are data for 28 x 100 mesh Ireland Mine coal preoxidized at low pressure (1.5 atm) during a previous experimental program.(22) Somewhat lower fluidity was obtained at much less severe conditions of preoxidation as noted below:

Pressure, atm	Preoxidation Temperature, °F	% Preoxidation
1.5	700	3.3
15	750	18.6

On increasing the severity of preoxidation to 13.8% at low pressure and at 700°F, the product showed no fluidity.

The Gieseler run on raw Illinois No. 6 coal showed a maximum fluidity of about 2.8. Only Run 5P1 preoxidized coal was tested, and this gave no fluidity reading at all. Runs 5P2 and 6P were not tested, but the results of a qualitative test discussed below indicated that they were equally inert.

b. Laboratory Fluidized Shock Heat Test

Another test was to heat the coal sample to 1500°F in about two minutes while fluidizing it in a stream of nitrogen. These tests gave the following results:

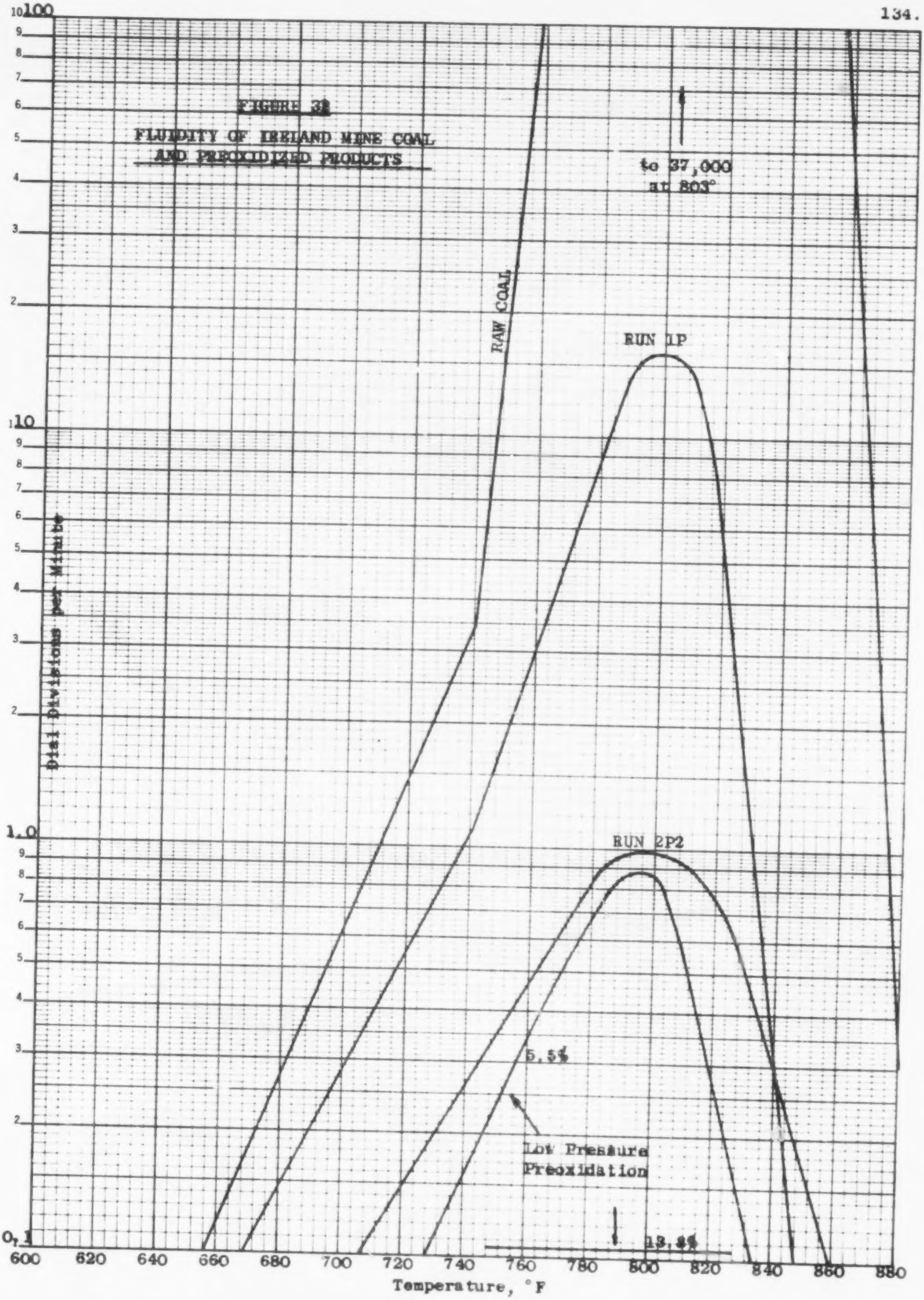


FIGURE 32

FLUIDITY OF IRELAND MINE COAL AND PREOXIDIZED PRODUCTS

K•E SEMI LOGARITHMIC 46 5493
3 CYCLES X 70 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

Product From Run No.	Temp., °F	% Preoxidation	Observation
4P	750	27.9	Slight agglomeration of particles.
5P1	800	11.9	Slight agglomeration of particles.
5P2	800	11.5	No agglomeration.
6P	810	8.7	No agglomeration.

The 1P and 2P2 products showed moderate to severe caking in the same test. It was concluded that preoxidized coals which showed no evidence of agglomeration would certainly pass through the gasifier without caking, while those which agglomerated slightly would probably make it. A coal which caked severely definitely would be inoperable.

c. Free Swelling Index

A third test which was tried was the Free Swelling Index (FSI). This test is primarily used to evaluate coking coals. The test consists of rapidly heating the coal to 1500°F in a special crucible and then examining the shape of the resulting pellet against a standard chart. We found that even a treated coal which had an FSI of only 1 (2P2 product) would agglomerate in the gasifier. The FSI method was considered inferior to the fluidized test and was dropped.

7. Probable Cause of Ineffective Preoxidation at Elevated Pressure

Our previous experience with preoxidation at atmospheric pressure predicts that the products from Runs 1P and 2P2 would have been completely operable in the gasifier. In addition, preoxidation of Ireland Mine coal at essentially Run 2P2 conditions, but at atmospheric pressure, provided an operable feedstock for the experimental work on IGT's HYGAS Process.(23)

Microscopic examination of the Runs 1P and 2P2 products showed no unusual features. In fact, the appearance of the particles was entirely similar to those produced during the laboratory study of preoxidation at atmospheric pressure, which are described in Section VI.

The Gieseler plastometer data for Ireland Mine coal preoxidized at low pressure would indicate both the Runs 1P and 2P2 products should have no fluidity.

A clue provided in Run 2P2 leads to a plausible explanation of the ineffectiveness of preoxidation at 15 atm. Previous data for preoxidation at atmospheric pressure(24) show that, at Run 2P2 conditions of temperature and severity of preoxidation, about 3 wt % of tar should have been produced. Such tars contain a large proportion of high boiling (+750°F) material. In Run 2P2 no heavy tar was produced. Rather, a light-yellow oil amounting to 0.20 wt % of the coal was collected with the condensed water. At 15 atm pressure, evaporation of the tarry matter from the surface of the coal particles is hindered. Thus, the material remains within the particles, encased in a rigid outer shell. When exposed to gasification conditions, this material breaks through the shell and causes caking or agglomeration.

C. Gasification

1. Introduction

At process conditions, there are four gaseous streams which enter at, or near, the bottom of the gasifier. The nature of these streams, as determined by heat and material balance calculations, made for Case I (Table 5), is shown below:

Mols/100 Pounds Dry Coal Feed to Preoxidizer

Stream No.	1 Preoxidizer Gas	2 Claus Tail Gas	3 Air	4 Steam	Total
CO_2	.188	1.946			2.134
N_2 + CO	1.330	4.329	4.353		10.012
H_2O	.558	.071		3.732	4.361
O_2	.076		1.157		1.233
	2.152	6.346	5.510	3.732	17.740

In the experimental work, the flows of CO_2, steam, coal and total inlet gas always were maintained in the proportions shown above. The tendency toward formation of local hot spots and, therefore, of ash slagging increases as the O_2 partial pressure at the entry point increases. The O_2 partial pressure can be varied, within limits, depending upon how the four inlet streams are combined:

Conditions at Feed Point of Preoxidized Coal	p_{O_2}, atm (system pressure = 15 atm)
Complete Mixing of all Streams	1.04
Streams 1 + 3 + 4	1.62
Streams 1 + 3	2.41
Stream 3 only	3.15

Other combinations are possible, but the above list serves to illustrate the range of partial pressures involved.

The gasifier was operated in the absence of acceptor since the phenomena of caking and ash slagging are independent of the presence of acceptor.

The two possible problems anticipated in operating the gasifier are caking and ash slagging. The first preoxidized feedstock, Run 1P product, was shown in the laboratory to be almost certain to cake in the gasifier. In order to determine the magnitude of the problem of ash slagging and to train the operators in running the gasifier, Disco char was used in the gasifier feedstock. The Disco char, made from Pittsburgh Seam coal, had been partially gasified at 1775°F in a previous experimental program, privately sponsored. Four runs were made with this material.

2. Operability with Respect to Ash Slagging

a. Runs OG2 and OG3

These runs were made with all the inlet streams fed to the bottom of the cone in the gasifier boot, as is shown in Configuration A, Figure 32. Since the two entry points are within one inch of each other, this configuration corresponds to perfect mixing of the four inlet gas streams. Run conditions and results are given in Table 35. In Run OG2 the air flow was adjusted to give an O_2 partial pressure of 0.43 atm at the bottom of the fluidized bed. Temperature profiles showed no evidence of ash slagging. After about nine hours of operation, the air flow (for Run OG3) was increased to give an inlet O_2 partial pressure of 1.04 atm. This air flow corresponds to the total air required at process conditions. Again, there was no evidence of ash slagging. After two hours, the run was ended and the vessel was disassembled for inspection. No trace of slag was found.

b. Runs OG4 and OG5

For these runs, the air and char entry point was raised, by means of a 1/4" OD tube, into the 4" diameter section of the gasifier bed, as was shown in Configuration B in the aforementioned figure. In Run OG4 the O_2 partial pressure at the inlet tube was 1.0 atm, corresponding to complete mixing of all process gas streams. After about nine hours of operation, which gave no evidence of slagging, the air flow (for Run OG5) was increased to give an O_2 partial pressure at the inlet tube of 2.4 atm, corresponding to mixing of Streams 1 and 3. Since the temperature profiles again showed no evidence of slagging, the run was ended after four hours. Inspection of the vessel showed that no slag had formed. The exit streams for these runs were not measured since the inlet flows were identical with Runs OG2 and OG3.

3. Operability with Respect to Caking - Use of Preoxidized Ireland Mine Coal

a. Run 1G3

This run was made at Run OG3 conditions, with all inlet streams fed to the bottom of the gasifier boot. Run conditions first were established by feeding Disco char to the gasifier. Preoxidized Ireland Mine coal from Run 1P was used in this run to provide a severe test of the concept of the use of an O_2-bearing carrier gas to prevent caking.

After 13 minutes of coal feeding, the pressure drop across the fluidized bed increased drastically, an indication that massive caking had occurred. Examination of the material recovered after disassembly of the vessel showed that the coal had melted with complete loss of particle identity and had cemented together the Disco char particles.

b. Run 2G3

This run was made at the same conditions as in Run OG3. Preoxidized Ireland Mine coal from Run 2P2 was used as feedstock. Almost immediately after feeding the coal, the product char contained numerous small

FIGURE 32

CONFIGURATION OF ENTRY POINTS IN GASIFIER

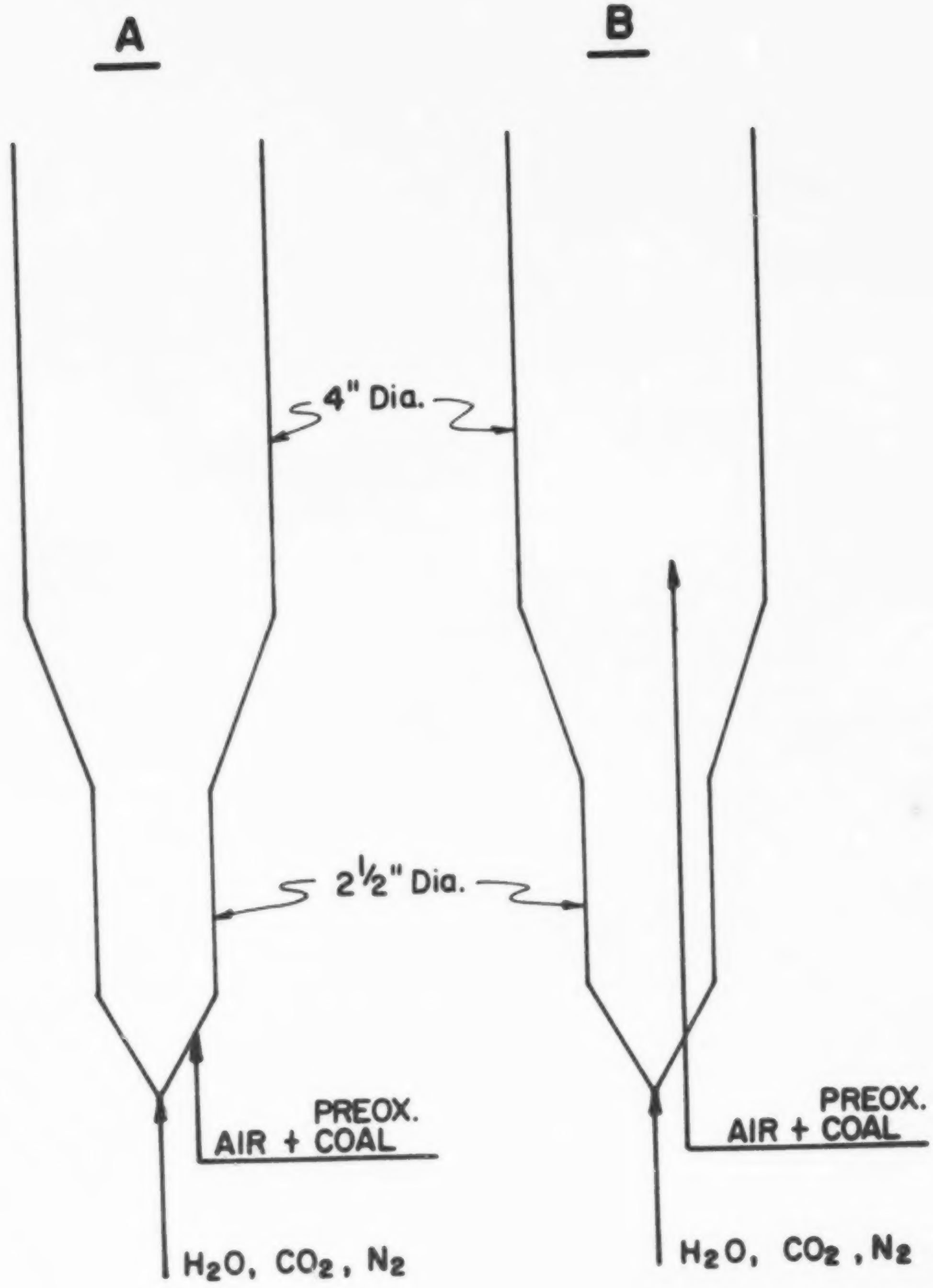

A
B
4" Dia.
2½" Dia.
PREOX.
AIR + COAL
H₂O, CO₂, N₂
PREOX.
AIR + COAL
H₂O, CO₂, N₂

TABLE 35

Conditions and Results for Gasification Runs

System Pressure: 15 atmospheres (206 psig)
Temperature : 1700° F

Run Number	OG2	OG3	5G3	6G3
Input				
Steam, SCFH	84	78	76	77
CO_2	38	35	38	38
N_2	64	64	64	64
Lift Gas				
Air, SCFH	43	105	105	105
N_2	99	34	34	34
Purges (N_2), SCFH to bed	13	13	5	5
Purges (N_2), SCFH above bed	18	24	6	9
Feedstock	28 x 100 Disco Char		Run 5P2	Run 6P
Feed Rate, lb/hr	4.93	5.50	4.15	4.83
Output				
Exit Gas Rate, SCFH (dry basis)	334	332	334	334
Composition, Mol %				
H_2	7.93	7.61	11.17	11.81
CH_4	.40	.39	.86	1.07
CO	9.90	10.33	13.81	13.78
CO_2	13.51	16.21	14.97	14.64
N_2 (by difference)	68.26	65.46	57.54	58.69
Outlet Gas, top of bed				
Composition, Mol %				
H_2O	14.60	13.68	15.03	15.58
H_2	7.04	6.82	9.67	10.23
CH_4	.36	.35	.75	.93
CO	8.78	9.27	11.96	11.94
CO_2	11.99	14.54	12.95	12.69
N_2	57.2	55.3	49.64	48.62
Flow Rate, SCFH, at top of bed	376	370	386	385
Fluidizing Velocity, ft/sec	.34	.33	.34	.34
Product Char, lb/hr	3.56	3.53	1.51	2.07
Duration of Coal Feeding, hr	9.5	2.0	8.0	7.6
Bed Weight, lb	7.10	7.02	4.56	4.14
Total Gasification Rate(1)	45	71	105	110
Percent Carbon Burnoff(2)	35	46	68	55

(1) Pounds Carbon gasified/min/lb Carbon in bed x 10^4.

(2) $\frac{100 \text{ (lb/hr Carbon gasified)}}{\text{(lb/hr Carbon in with coal)}}$

agglomerates. The amount of agglomerates continued to increase and the run was ended after 45 minutes. After shutdown, the bed material drained freely from the gasifier. Microscopic examination of this material showed that the coal particles had not lost their identity, but that they had adhered to the Disco char particles, and to themselves.

c. Run 4G3

This run was made at Run OG3 conditions using the Run 4P product as gasifier feedstock. The history of this feedstock is summarized below:

	Temp., °F		Percent Preoxidation
1st Stage (Run 2P2)	750		18.6
2nd Stage (Run 4P)	750		9.3
		Total	27.9

After 2.5 hours of coal feeding, steady state had been reached with respect to displacement of the start-up bed inventory. At that time a severe leak developed in the char outlet line, which forced a shutdown before any measurements were made on the exit streams.

Examination of the reactor after disassembly showed no trace of ash slagging or caking/agglomeration.

4. Operability During Gasification - Use of Preoxidized Illinois No. 6 Coal

The preoxidized coal products from Runs 5P2 and 6P were fed to the gasifier operated at full process air flow in Runs 5G3 and 6G3, respectively. Run conditions and results are summarized in Table 35, and some properties of the char products are shown in Table 36.

Both runs were completely operable, with no trace of agglomeration/caking or of ash slagging.

5. Conclusions

Results indicate that ash slagging in the gasifier does not pose a problem with either Pittsburgh Seam or Illinois No. 6 coals.

Preoxidized Illinois No. 6 coal and preoxidized Pittsburgh seam coal have both been fed to the gasifier without caking. As far as the gasifier vessel is concerned, these materials represent satisfactory noncaking feeds. However, the amount of preoxidation required for the Pittsburgh Seam coal is more than three times that which corresponds to adiabatic conditions in the process preoxidizer.

D. Integrated Operation with Acceptor Circulation

1. Introduction

A "demonstration" run was scheduled in order to give an appreciation of possible problem areas. The purposes of the run were:

TABLE 36

Properties of Feeds and Products - Gasification Runs

Run Number	5G3		6G3	
	Preoxidized Coal Feed	Char Product	Preoxidized Coal Feed	Char Product
Hydrogen, Wt % (dry basis)	3.29	.46	3.44	.50
Carbon	67.02	65.64	70.61	68.94
Nitrogen	1.44	.39	1.36	.55
Oxygen (by difference)	7.98	-.10	5.80	-1.09
Sulfur	3.72	2.47	3.40	2.22
Ash	16.55	31.14	15.39	28.88
Sulfide Sulfur	.01	1.73	.03	1.31
Pyritic Sulfur	1.11	0	1.37	0
Sulfate Sulfur	.27	0	.07	0
Mean Particle Diameter, inch	.0171	.0155	.0175	.0155
Mean Particle Density, lb/ft^3	49.8	34.1	51.8	43.2
Ash Composition, Sulfur-free, Wt %				
Al_2O_3	18.5			
SiO_2	53.6			
Fe_2O_3	17.4			
CaO	3.7			
MgO	1.1			
Na_2O	1.2			
K_2O	2.4			
TiO_2	0.9			
P_2O_5	0.3			
Total	99.1			

a. To determine whether or not kinetic limitations exist on:

  (1) Rate of desulfurization of the char feed in the gasifier at process conditions.

  (2) Rate of pickup of H_2S by the acceptor in the gasifier.

b. To determine magnitude of acceptor activity loss at process conditions.

c. To determine ash slagging characteristics of fuel char in the regenerator.

d. To determine the extent and nature of any deposits formed in the gasifier or regenerator.

e. To obtain semi-quantitative data on rate of gasification.

No air was used in the run in order to simulate the reducing conditions in the upper part of the process gasifier where most of the char desulfurization and H_2S pickup by the acceptor occurs. Added H_2 and recycle gas were used to control the partial pressures. Process heat was supplied via the electrical heaters and by the acceptor reaction. Ungasified Disco char, made from Pittsburgh Seam coal, was used as gasifier feedstock in order to avoid any difficulty caused by caking. The regenerator fuel char was the same material which had been partially gasified during break-in operations.

Regeneration of CaS was in the manner of Case I in the feasibility study, except that by use of recycle gas, the ratio of exit gas/total sulfur in the regenerator was high enough to avoid any equilibrium limitations on SO_2 partial pressure. (Study of the full-blown Case I regenerator conditions is a separate experimental program and will be undertaken during Phase II-B of the contract).

2. Run D1B

During the run, thirteen calcining-recarbonation cycles were achieved.

A major purpose of the demonstration run was to sample the gasifier product gas to determine the extent of desulfurization. During the fifth cycle, an attempt to obtain a gas sample was thwarted by a corroded sample valve. At this point, a voluntary shutdown was made in order to replace the valve. The opportunity was taken to dismantle the vessels for inspection. No evidence of any deposits or of ash slagging in the regenerator was found.

Due to inexperience in operating the regenerator, the electrical power input, used to compensate for heat losses, was too low. As a result the bed temperature tended to level out at 1840°F rather than at the process temperature of 1880°F. During the first five cycles, the bed temperature purposefully was held at 1840°F in order to reduce the chance of early termination caused by ash slagging.

On resumption of the run the bed temperature was increased to 1860°F since no slagging had occurred earlier. Again, the bed temperature was kept lower than the process temperature in order to obtain the desired data on the acceptor sulfur cycle, uncomplicated by possible ash slagging.

Acceptor circulation was continued until the thirteenth cycle when the regenerator thermowell ruptured. An immediate shutdown was required. The thermowell, made of Hastelloy X, must withstand the full system pressure of 206 psig. Failure was caused by severe, localized, corrosion by sulfur. In future work the thin-walled Hastelloy X thermowell will be replaced by a larger diameter thick-walled thermowell made of Type 310 stainless steel, which has better resistance to sulfur corrosion.

Inspection of the vessels after the final shutdown again showed no evidence of deposits or ash slagging in the regenerator.

Summarized run conditions and results are shown in Table 37 for the gasifier and regenerator. Results are shown for the tenth calcining-recarbonation cycle, at which time the product gas sample was taken. Analyses of the char feeds and products are given in Table 38.

3. The Sulfur Cycle

a. Acceptor Sulfur Cycle

A discussion of the sulfur reactions which are involved is given in Section III-B-2 (feasibility study).

The sulfur contents of the recirculating acceptor, sampled periodically during the run, are shown below:

Cycle No.	4	5	7	10	13
Gasifier Samples					
% CaS content*	9.0	11.2	10.9	9.5	5.2
Regenerator Samples					
% CaS content	ND	9.8	ND	6.4	3.6

* mol % of total calcium in form of CaS.
ND not determined.

During the first part of the run (first five cycles), sulfur accumulated in the acceptor. This is attributed to the low (1840°F) regenerator temperature which suppressed the rejection of sulfur by the reaction,

$$3/4\ CaSO_4 + 1/4\ CaS = CaO + SO_2 \qquad (3)$$

During the second part of the run, the increased bed temperature (1860°F) allowed the accumulated sulfur inventory to be depleted. During the tenth cycle the data in Table 37 show that the SO_2 partial pressure in the exit gas was 0.035 atm. The equilibrium partial pressure (Figure C-3) for reaction (3) at 1860°F is 0.070. Since this cycle occurred during the period of depletion, the resulting ΔP_{SO_2} driving force of 0.035 atm thus is more than adequate to reject sulfur by reaction (3).

TABLE 37

Conditions and Results for Demonstration Run - Gasifier

System Pressure: 15 atmospheres (206 psig)
Temperature: 1700° F

Run Number	D1B
Feedstock	28 x 100 mesh fresh Disco Char
Acceptor	16 x 28 mesh Tymochtee 9 dolomite
Cycle Number	10
Acceptor Circulation rate, lb/hr ($MgCO_3 \cdot CaCO_3$ basis)	12.8
Acceptor Activity	.573 = mol ratio, $\frac{CaCO_3}{CaCO_3 + CaO}$
Char Feed Rate, lb/hr (dry basis)	4.54
Input	
Steam, SCFH (to boot)	73
CO_2 (to boot)	53
N_2 (to boot)	53
H_2 (to boot)	15
Recycle Gas, SCFH	
Fluidizing, to boot	73
Char lift line, above boot	47
Purges (N_2), SCFH to bed	10
Purges (N_2), SCFH above bed	32
Output	
Exit Gas Rate, SCFH (dry basis)	324
Composition, Mol %	
H_2	17.86
CH_4	1.59
CO	19.40
CO_2	14.96
N_2 (by diff.)	46.19
Outlet Gas, Top of Bed	
Composition, Mol %	
H_2O	17.97
H_2	16.25
CH_4	1.45
CO	17.65
CO_2	13.61
N_2	33.06
Flow Rate, SCFH, at top of bed	358
Product Char, lb/hr	3.12
CO_2 to Acceptor, SCFH	13.1
Fluidizing Velocity, ft/sec	
top of bed	.32
boot (acceptor layer)	.64
Duration of Acceptor Circulation, hr	9.6
Bed Weight, lb	7.58
Total Gasification Rate(1)	31
% Carbon Burnoff(2)	31

(1) Lbs. Carbon gasified/min/lb Carbon in bed x 10^4

(2) $\frac{100\ (\text{lb/hr Carbon gasified})}{(\text{lb/hr Carbon in with Char})}$

TABLE 37 - Cont'd.

Conditions and Results for Demonstration Run - Regenerator

System Pressure: 15 atmospheres (206 psig)
Temperature : 1860° F

Run Number	D1B
Fuel Char	28 x 100 mesh partially gasified Disco Char
Acceptor	16 x 28 mesh Tymochtee 9 dolomite
Cycle Number	10
Acceptor Circulation Rate, lb/hr ($MgCO_3 \cdot CaCO_3$ basis)	12.8
Acceptor Circulation Rate, lb/hr (as fed basis)	8.50
Acceptor Activity	.573
Acceptor Sulfur Content, Wt % (as fed basis)	2.10
Fuel Char Feed Rate, lb/hr (dry basis)	1.40
Input	
Recycle, SCFH	
Fluidizing	153
Char and Acceptor Lift Line	242
Air	115
N_2	85
CO_2	8
Purges (N_2), SCFH to bed	5
Purges (N_2), SCFH above bed	19
CO_2 from Acceptor, SCFH	13.1
Output	
Exit Gas Rate, SCFH (dry basis)	244
Composition, Mol %	
CO_2	17.01
CO	.80
SO_2	.23
N_2 (by diff.)	81.96
Flow Rate, SCFH, at top of bed	635
Fluidizing Velocity, ft/sec.	1.04
Acceptor, lb/hr ($MgO \cdot CaO$ basis)	7.02
Acceptor, Sulfur Content, Wt %	1.74
Overhead Ash, lb/hr	.652
Partial Pressure Driving Forces, atmospheres	
ΔP CO_2 Bottom of Bed(1)	2.74
ΔP CO_2 Top of Bed(1)	2.05
ΔP SO_2 Top of Bed(2)	.035
Inlet O_2 Partial Pressure, atmospheres	.58

(1) Below equilibrium partial pressure in the reaction:

$$CaCO_3 = CaO + CO_2$$

(2) Below equilibrium partial pressure in the reaction:

$$3/4\ CaSO_4 + 1/4\ CaS = CaO + SO_2$$

TABLE 38

Analyses of Char Feeds and Product
Run D1B

	Gasifier		Regenerator	
	Feedstock	Product	Fuel Char	Overhead Ash
Hydrogen, Wt % dry basis	2.63	0.52	0.69	0.3
Carbon	76.24	75.47	76.98	50.2
Nitrogen	1.77	.065	0.91	~ 0
Oxygen (diff.)	4.66	0.28	1.72	~ 0
Sulfur	1.27	0.45	0.84	2.54
Ash	13.43	23.22	18.86	46.9
Sulfide Sulfur, Wt %	0	0.35	0.09	0.46

Ash Composition, Sulfur-free basis

Al_2O_3, Wt %	23.2
SiO_2	51.6
Fe_2O_3	13.1
CaO	4.7
MgO	0.8
Na_2O	0.4
K_2O	1.8
TiO_2	1.0
P_2O_5	0.4
Total	97.0

The H_2S content of the gasifier product dry gas was below the detection limit of 30 ppm of the gas chromatographic method used for analysis. As shown in the feasibility study, the equilibrium content of the wet gas at process conditions is 230 ppm. Undoubtedly, most of the H_2S content of the product gas was absorbed by the ammonia-rich condensate formed by the unreacted steam and the ammonia generated from the char nitrogen in the gasifier. The equilibrium H_2S content of the gasifier wet product gas, calculated for the data in Table 37 and the equilibrium for the reaction, $CaO + H_2S = CaS + H_2O$ (shown in Figure C-2), corresponds to 4.5 gm/hr of sulfur. Analysis of the recovered condensate drained from the catchpots below the exit gas coolers, C1 and C4 (Figure 12), showed that the absorbed H_2S, present as NH_4HS in the condensate, amounted to only 2.2 gm/hr sulfur, or roughly one-half of the minimum possible amount of H_2S in the wet gas. Some of the "missing" H_2S flashed off the condensate as it was being drained from the catchpots at 15 atm pressure into receivers at room conditions.

In future work a small slipstream of the hot, wet gas taken downstream of the gasifier solids filters will be throttled to atmospheric pressure, cooled and then passed through aqueous dilute mineral acid to insure that all the H_2S is swept into a suitable absorber for subsequent analysis. Back titration of the dilute acid also will give the total ammonia content of the wet gas.

b. Char Desulfurization in the Gasifier

Sulfur balances based on data in Tables 37 and 38 show that 76% of the feed sulfur was removed in the gasifier, based on the total sulfur content of the product char. However, most of the residual sulfur was in the form of CaS in the char ash, as is shown by the sulfide sulfur analysis in Table 38. From a process standpoint, it is immaterial whether the sulfur content of the gasifier feedstock leaves the gasifier as CaS in the acceptor or as CaS in the char ash. After correction for the CaS content of the product char, the removal of organic sulfur* was 93%.

A high degree of sulfur removal also occurred in Runs 5G3 and 6G3, made with preoxidized Illinois No. 6 coal feedstock in the absence of acceptor. From Tables 35 and 36, the extent of sulfur removal based on total sulfur content and organic sulfur content have been calculated as follows:

	Percent Sulfur Removed from Feedstock	
Run No.	Total	Organic
5G3	76	91
6G3	72	86

Comparison of the outlet gas compositions for these runs with equilibria in the Fe-O-H-C-S system shows that the FeS and FeO are the stable phases at the conditions used and that sulfide sulfur is formed by the reaction,

$$FeO + H_2S = FeS + H_2O.$$

---

* Organic sulfur defined as: Total sulfur minus sulfide sulfur.

c. Sulfur Balances

As discussed above, the H_2S content of the gasifier product gas could not be determined because of loss from the condensate. Furthermore, the accumulation-depletion of CaS "flywheel" in the acceptor inventory somewhat obscures the true behavior of the acceptor sulfur cycle. The following approach was used in order to present the sulfur balances: (1) In the gasifier a hypothetical, sulfur-free acceptor was assumed to enter from the regenerator and the H_2S content of the wet product gas was assumed to be at equilibrium. The sulfur content of the acceptor leaving the gasifier then was obtained by difference. (2) In the regenerator, measured values for the output streams were used, along with the aforementioned sulfur content of the gasifier acceptor stream. (3) The excess sulfur leaving the regenerator then was rationalized with the rate of depletion of the "flywheel". The calculations were based on data in Tables 37 and 38. Three additional measurements on minor streams were made, but are not included in the tables:

(1) COS content of gasifier dry product gas.
(2) SO_2 content of condensate from regenerator gas cooler, C-6.
(3) Sulfur content of regenerator back-up filter.

Results of the calculations are shown in Table 39.

The by-difference value of the sulfur content of the acceptor leaving the gasifier corresponds to 1.8 mol % conversion of the total calcium to CaS, the same conversion as used in the feasibility study (Table 5). The regenerator balance shows considerably more sulfur out than in. This excess sulfur is rationalized with the depletion data shown on page 139 at the bottom of Table 39. As shown, the agreement is good. As can be seen in Table 39, the amount of sulfur in the "flywheel" is large compared with the total amount of sulfur entering the system.

In future studies on acceptor activity with respect to the sulfur and CO_2 reactions, the temperature and sulfur/gas ratio in the regenerator will be adjusted to avoid the accumulation of sulfur in the recirculating acceptor.

d. Elemental Sulfur Content of the Regenerator Offgas

If equilibrium is assumed for the reaction,

$$2\ CO + SO_2 = 1/2\ S_2 + 2\ CO_2, \qquad (4)$$

the partial pressure of elemental sulfur in the offgas can be calculated from the data in Tables 37 and C-3. The result is 0.00473 atm at 1860°F, which corresponds to 5.9 gm/hr sulfur. At the end of Run D1B, only a small amount of elemental sulfur corresponding to about 0.1 gm/hr was found. Its location was in the piping between the back-up filter and the gas cooler C6 (Figure 12).

TABLE 39

Sulfur Balance - Run D1B
Cycle Number 10

	Grams/hour Sulfur
Gasifier	
In	
Feed Char	26.2
Acceptor *	0
Out	
Product Char	6.4
H_2S to Gas **	4.5
COS	0.2
to Acceptor (by diff.)	15.1
	26.2
Regenerator	
In	
Acceptor	15.1
Fuel Char	5.3
	20.4
Out	
SO_2 in Gas	21.1
SO_2 in Condensate	0.1
Overhead Ash	7.5
Backup Filter	0.9
	29.6
	excess sulfur = 29.6 - 20.4 = $\boxed{9.1}$
Depletion of Acceptor Sulfur Inventory	
Fed to Regenerator	81.0
Less Sulfur Added in Gasifier	15.1
	65.9
Out from Regenerator	55.5
	excess sulfur = 65.9 - 55.5 = $\boxed{10.4}$

* By assumption.

** Based on equilibrium content.

The success on the Case I sulfur rejection method described in Section III-B-2 (feasibility study) depends upon close approach to equilibrium in reaction (4). If this situation occurred in Run D1B, then about 5.8 gm/hr of elemental sulfur must be accounted for. As shown in Table 38, the sulfur content of the overhead ash material is about three times greater than that of the fuel char. At combustion conditions, the fuel char sulfur can be expected to form SO_2 rather than to remain in the overhead ash. As shown in Table 39, the sulfur content of the overhead ash material corresponds to 7.5 gm/hr sulfur which, after correction for the sulfide sulfur content,* is sufficient to account for the necessary elemental sulfur.

In the continuous gasification unit, the regenerator offgas which contains the overhead ash material has been cooled to about 300°F before reaching the external cyclone. This temperature is favorable for condensation of sulfur vapor and subsequent absorption by the high surface area ash material.

A single batch extraction of the overhead ash with boiling toluene, followed by evaporation of the solvent, produced elemental sulfur crystals. Qualitatively, the amount of sulfur recovered appeared to be too small to account for the measured sulfur content of the overhead ash. In future work the overhead ash will be extracted exhaustively in a modified Soxhlet apparatus to determine whether absorbed sulfur can account for all of the observed increase in sulfur content.

An alternate explanation of the enhanced sulfur content is that organic C-S bonds are formed from the elemental sulfur and the unburned carbon at regenerator conditions.

4. Acceptor Activity

A major purpose of Run D1B was to determine the effects on acceptor activity toward the CO_2 acceptor reaction of exposure to several cycles of calcining and recarbonation.

During the run, the circulating acceptor inventory was sampled periodically, and its $CaCO_3$ content was determined. The acceptor activity is defined as the mol ratio,

$$\text{Activity} = \frac{CaCO_3}{CaCO_3 + CaO}$$

For the acceptor samples withdrawn from the gasifier "undercarbonation", never was observed, i.e., the carbonation ratio always was consistent with the intrinsic activity, as was shown by the fact that no weight gain occurred on prolonged exposure to CO_2 at conditions which were kinetically very favorable for the reaction, $CaO + CO_2 = CaCO_3$.

Acceptor samples withdrawn from the regenerator always were completely calcined, i.e., any $CaCO_3$ present was undetectable.

No acceptor makeup was used during the run. Therefore, the acceptor activity determined for each sample could be associated with the completion of a definite number of calcining-recarbonation cycles. A cycle is completed when the number of mols of CaO fed from the L-4 acceptor feed hoppers equals the number of mols of CaO contained inside the regenerator, gasifier, and the standlegs. Results are shown below:

Cycle	Activity
4	0.68
5	0.65
7	0.66
10	0.57
13	0.55

An accurate estimate of the equilibrium activity of the recirculating CO_2 acceptor at the conditions of Run D1B would require the completion of about 40 cycles, based on our previous work for the Office of Coal Research.$^{(1c)}$ However, comparison of the above results with the OCR data indicates that the equilibrium activity will be higher than that estimated in the feasibility study (page 21 of this report).

5. Char Combustion in the Regenerator

A characteristic of the run was that the CO content of the regenerator offgas could not be increased beyond 0.8 mol %. Attempts to increase the CO content by temporarily decreasing the air/fuel char ratio were without effect, showing that the low reactivity of Disco char toward the reaction, $C + CO_2 = 2\ CO$, which must occur in the upper part of the bed above the combustion zone, was responsible. It is noteworthy that no CaS-$CaSO_4$ transient liquid was observed at this low level of CO content. Previous work with low-rank Western coals, which have a high lime content in the ash, showed that maintenance of 2-3% CO in the regenerator offgas is necessary to suppress formation of the transient liquid in the coal ash.

The carbon burnout of the regenerator fuel, calculated by carbon and ash balances based on data in Tables 37 and 38, was 74% vs 90% which was assumed for the feasibility study. The observed lower burnout is consistent with the low reactivity of Disco char which led to the low CO level mentioned above. In future work the bed temperature will be raised above 1860°F to determine the limit dictated by ash slagging. Higher temperatures will improve the burnout situation.

6. Nitrogen Removal from the Gasifier Feedstock

Removal of nitrogen from the feed char, calculated from data in Tables 37 and 38 was nearly complete at 97%. Analysis of the recovered condensate showed that at least 30% of the incoming char nitrogen had been converted to ammonia.

7. Kinetics of the Gasification Reactions

The five gasification runs made so far were concerned with potential operability problems. Gasification rate data obtained during the runs merely were byproducts. As shown in Tables 35 and 37, the rates range from 110 x 10^{-4} lb carbon gasified/lb carbon in bed/min at bottom-of-gasifier conditions to 31 x 10^{-4} at top-of-gasifier conditions, compared with the overall rate of 75 x 10^{-4} which was assumed in the feasibility study.

Although it seems possible that the assumed rate can be achieved, a systematic study will be necessary in which a preoxidized coal feedstock first will be processed at bottom-of-gasifier conditions. The char product will then be used sequentially as feed char to succeeding runs in which the partial pressures of reactants and products will approach those at top-of-gasifier conditions.

VI. LABORATORY STUDY OF COAL PREOXIDATION KINETICS

A. Introduction

The objective of the study was to determine the rate of reaction of oxygen with Ireland Mine coal, as influenced by the following variables:

1. temperature
2. O_2 partial pressure
3. extent of preoxidation
4. coal particle size

Conditions and results for 17 runs are summarized in Table 40.

TABLE 40

Summary of Results
Laboratory Study of Kinetics of Coal Preoxidation
1 atm system pressure

Run No. 1363-	°F	Mol % O_2 in inlet gas	Grams Dry Coal	Duration Minutes	CO Removal Method	Total % Preoxidation (1)	Total % Preoxidation (2)	% H Removed (3)	Wt % Loss	Particle Density (4)
				28 x 35 mesh coal						
39	650	30.1	4.87	24.0	Hopcalite	11.4	11.8	17	4.5	77.7
35	↓	66.8	5.16	21.5		17.7	--	25	3.3	79.3
38		90.0	4.84	24.0	↓	20.6	--	28	3.7	80.9
32	700	30.1	3.95	21.0		19.2	--	28	5.9	63.3
30	↓	66.8	5.15	21.5		26.0	--	31	7.3	70.0
29		90.0	5.28	21.5		--	--	--	7.4	75.5
36	750	30.1	3.94	17.5		26.4	--	28	11.3	53.6
39	↓	66.8	5.01	21.0		45.8	--	42	17.8	63.2
37		90.0	3.98	17.5		41.2	--	40	14.5	65.0
44	750	30.1	2.31	14.5	cuprous sulfate	24.4	24.1	27	12.9	
41	↓	66.8	1.96	15.0		33.3	34.6	35	14.1	
40		90.0	1.99	16.0	↓	43.2	--	45	13.6	
				65 x 100 mesh coal						
20	600	66.8	4.82	19.0	cuprous sulfate	18.1	19.3	28	0.0	
24	658	30.1	5.00	21.0		19.1	24.1	22	4.7	86.0
19	730	66.8	5.12	19.0	↓	41.4	35.5	46	11.5	86.8
26	650	66.8	5.06	20.0	Hopcalite	28.3	--	36	5.3	
27	660	90.0	3.04	21.0		33.9	--	41	8.3	

(1) 100 x grams O_2 consumed/gram dry coal, from wt gain of absorbers.
(2) 100 x grams O_2 consumed/gram dry coal, from exit gas analysis.
(3) Wt % coal hydrogen reacted to form water.
(4) Lb/ft^3, measured in mercury at 1 atmosphere.

B. Experimental

A schematic diagram of the apparatus is shown in Figure 33. The reactor was a 1" diameter quartz tube containing 2 to 5 grams of coal and 50 grams of 200 x 325 mesh fused periclase. The periclase acted as a heat sink and diluent, and thereby allowed good temperature control during the highly exothermic preoxidation reactions. The system was operated at atmospheric pressure. All runs were made with Ireland Mine coal.

The inlet gas was an O_2-He mixture of known composition, at a flow rate of 1.50 SCFH. The gas entered the top of the reactor, passed to the bottom through an axial diptube, reversed direction, and fluidized the coal and periclase. The reactor was heated by being immersed in an air-fluidized bed of silica sand which in turn was heated by an external electric resistance furnace. Reaction temperature was measured by a calibrated thermocouple immersed in the fluidized bed of coal-periclase and was recorded continuously. The recorder also controlled the net rate of heat input to the fluidized sand bath furnace to hold the reaction temperature to within 3°F of the desired value. The product gas passed through a series of absorbers which removed the reaction products, as follows:

Product	Absorbent
H_2O	magnesium perchlorate
CO_2	Ascarite
CO	cuprous sulfate

The residual gas was then substantially a binary mixture of O_2 and He. Its composition was monitored continuously by a thermal conductivity cell in which the reference gas was the original O_2-He mixture used as the reactor inlet gas. The cell was calibrated, for each particular inlet gas, by passing known O_2-He mixtures through it. The method of calibration involves the use of a mathematical model of the thermal conductivity cell which was developed during previous work for the Office of Coal Research (reference 1c, Book 2). With this method of analysis for O_2, a difference in concentration of 0.01 mol % between the residual and inlet gas easily can be measured.

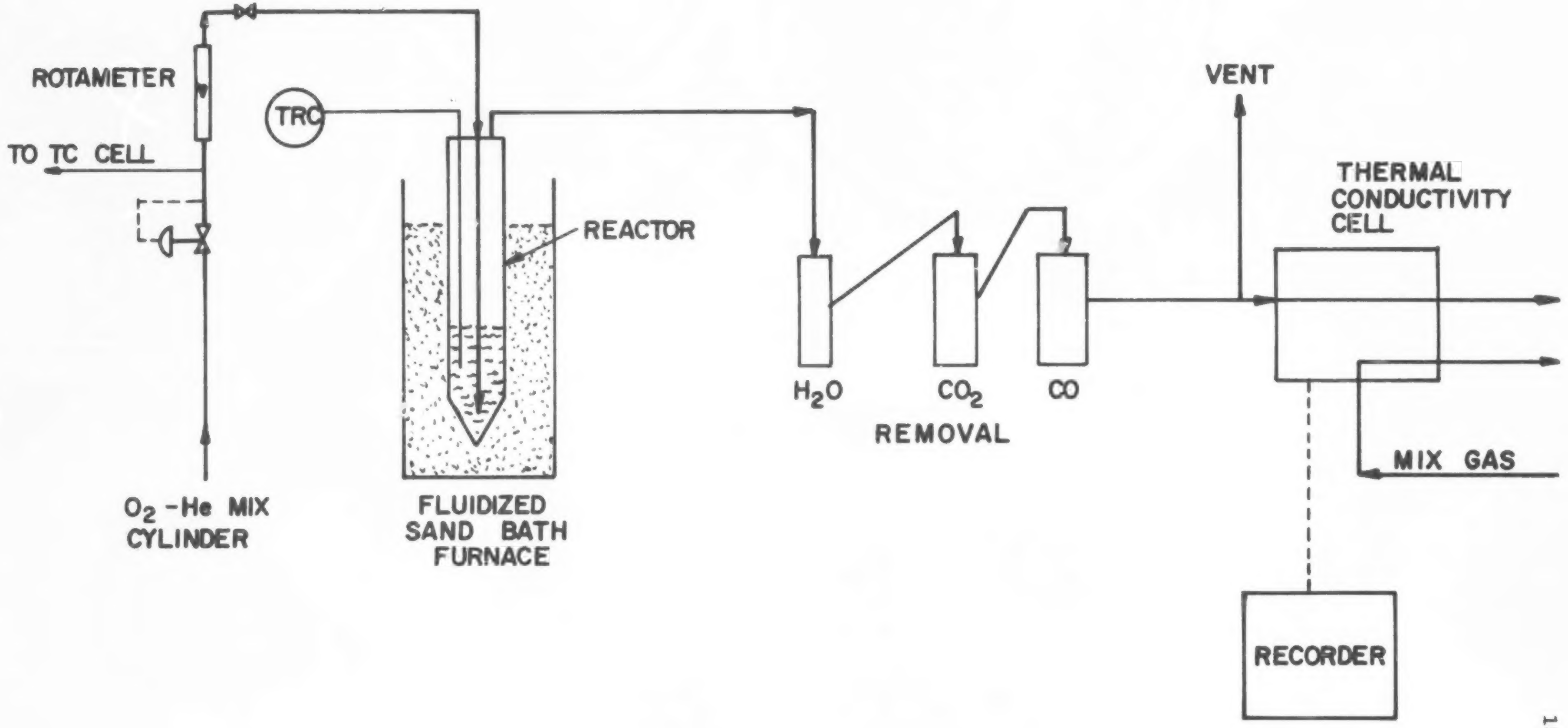
After assembly of the apparatus, inlet gas was passed through the system with the reactor at room temperature in order to establish the recorder baseline of the thermal conductivity cell output. The fluidized sand bath furnace, which was mounted on pulleys, then was raised quickly to surround the reactor. The coal-periclase bed reached the desired reaction temperature in less than two minutes. After about 20 minutes, the fluidized sand bath furnace was lowered and the reactor was quenched to room temperature by spraying it with water.

The rate of O_2 consumption at any time was calculated in a straightforward manner, based on the known flow of O_2 content of the inlet gas and the measured O_2 content of the residual gas. The cumulative extent of preoxidation up to any time was determined by numerical integration of the rate versus time data.

The ratio of coal to inlet O_2 was chosen to give a low conversion of the O_2, i.e., differential rate data with respect to O_2 partial pressure were obtained.

FIGURE 33

SCHEMATIC DIAGRAM
APPARATUS FOR STUDY OF COAL PREOXIDATION KINETICS


ROTAMETER
TRC
TO TC CELL
REACTOR
O_2 -He MIX CYLINDER
FLUIDIZED SAND BATH FURNACE
H_2O
CO_2
CO
REMOVAL
VENT
THERMAL CONDUCTIVITY CELL
MIX GAS
RECORDER

C. Rate of Reaction of Oxygen with Coal

The desired rate data were obtained reliably for only six runs. The data are plotted in Figure 34 as percent preoxidation per minute vs. cumulative percent preoxidation, where percent preoxidation is defined as 100(grams O_2 consumed/gram dry coal). The data also are shown in Figure 35 as cumulative percent preoxidation vs. time.

Use of cuprous sulfate (Cosorbent, Burrell Corp.) as the CO absorber led to difficulties because breakthrough occurred in an unpredictable manner. We attempted to alleviate the situation by replacing the cuprous sulfate with an active cupric oxide catalyst (Hopcalite) which oxidized the CO. We believed that the effect of the reaction, $CO + 1/2\ O_2 = CO_2$, on the response of the differential thermal conductivity cell, which was used to measure the O_2 content of the product gas, could have been neglected. However, when the run data were worked up, comparison of the total O_2 consumed, as determined by the product gas composition, with that determined by the weight gain of the CO_2 and H_2O absorbers showed that, in general, the use of Hopcalite gave rates that were too low. The effect of CO oxidation was small only in Run 39, made at the lowest conditions of temperature and O_2 inlet concentration.

The weight gain of the CO_2 and H_2O absorbers represents most of the O_2 consumed since previous work has shown that only about 6-8% of the O_2 consumed appears as CO over a wide range of temperatures and inlet O_2 concentrations.

Subsequently, the cuprous sulfate method was tried again with smaller coal charges and with two absorbers in series. No breakthrough occurred and the method now appears to be reliable. Lack of time prevented repetition of the Hopcalite runs.

1. Effect of Temperature

Only two pairs of runs, 39-44 and 20-19 (Figure 34) shared common conditions of inlet O_2 concentrations and particle diameter. Characteristically, the decline in rate with increasing extent of preoxidation is less severe at the higher temperatures.

2. Effect of Inlet O_2 Concentration

Only one pair of runs, 44-41, shared common conditions of temperature and particle diameter. On increasing the O_2 concentration by a factor of 2.2, the rate increased by a factor of 1.5 to 1.7, depending upon the extent of preoxidation.

3. Effect of Extent of Preoxidation

Characteristically, all the curves in Figure 34 show a high initial rate of reaction which is compatible with the heterogeneous nature of bituminous coal. The most reactive portion of the coal preferentially forms water, rather than carbon oxides, as determined qualitatively by observation of the rate of appearance of condensed water in the tube connecting the reactor and the H_2O absorber during the first few minutes of each run.

FIGURE 34

REACTION RATE VS. CUMULATIVE PERCENT PREOXIDATION

1 ATMOSPHERE PRESSURE

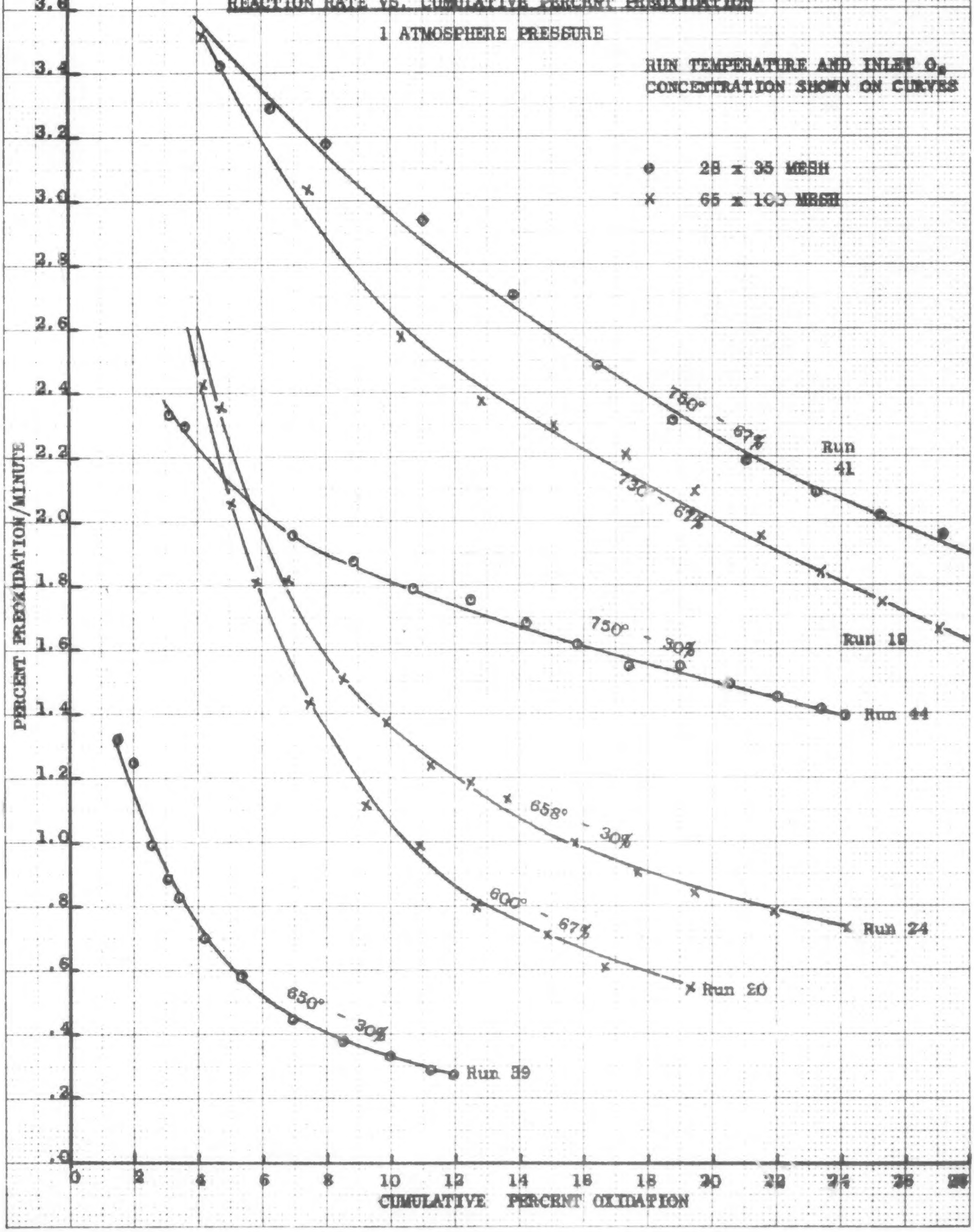


FIGURE 35

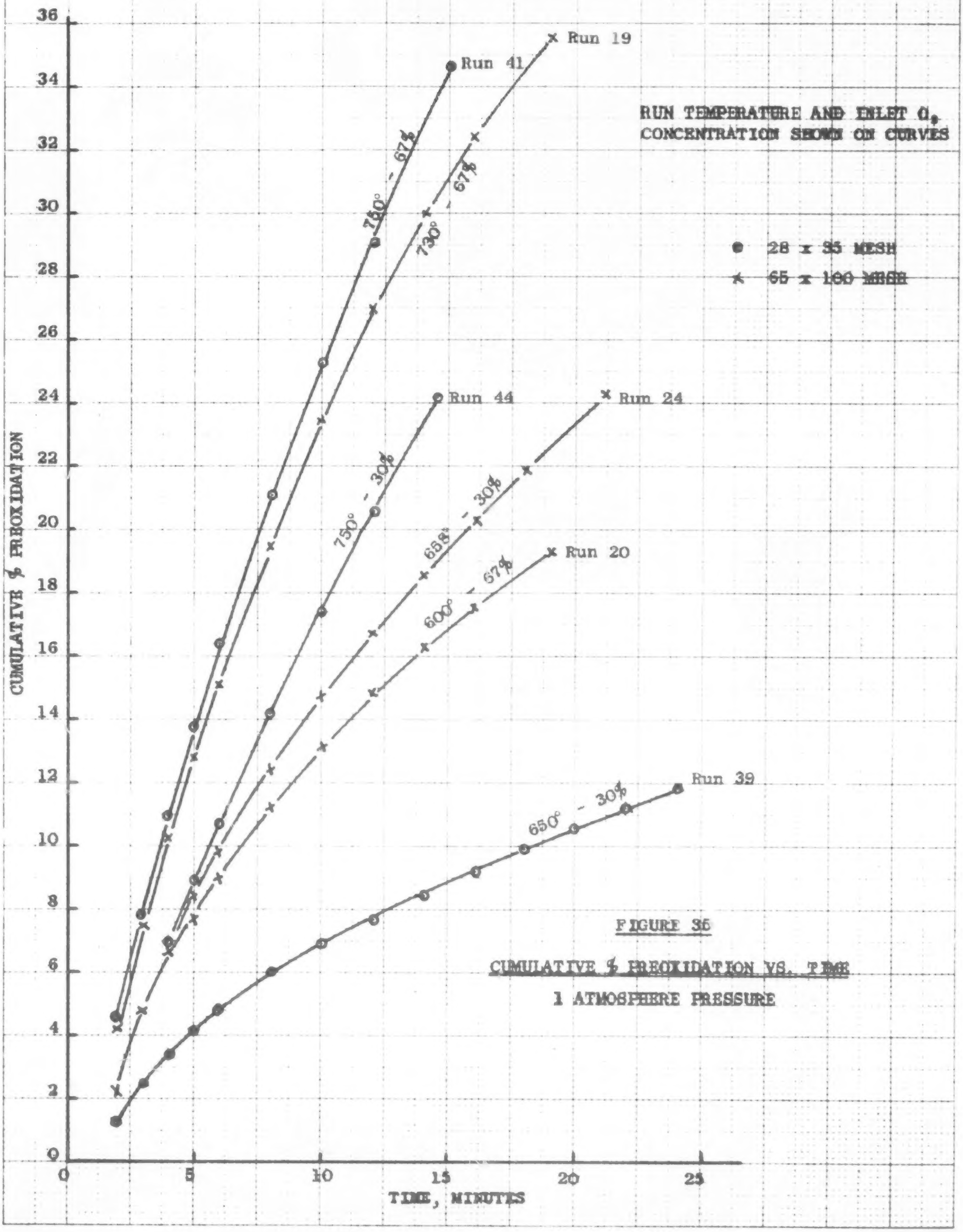


FIGURE 35

CUMULATIVE % PREOXIDATION VS. TIME

1 ATMOSPHERE PRESSURE

4. Effect of Particle Diameter on Rate of Reaction with O_2

At the lower temperature level (~ 650°F), comparison of Runs 39 (28 x 35 mesh) and 24 (65 x 100 mesh) shows that the rates for the smaller particles were higher by a factor of about 4.0, depending somewhat on the level of preoxidation.

If the rate of reaction were proportional to the external surface area of the particles, then the rate should have been higher by a factor of 8.0 for the smaller particles. If the rate of reaction were proportional to the particle diameter (shrinking core model with reaction rate controlling), then the rate should have been higher by a factor of 2.8. The observed intermediate value indicates that the mechanism of preoxidation probably involves the external surface area. The particle density (Figure 37) and microscopic examination both showed that some swelling of the 28 x 35 mesh particles had occurred.

At the higher temperature level (~ 750°F), the rates for Runs 19* (65 x 100) mesh and 41 (28 x 35 mesh) were substantially identical over the entire range of extent of preoxidation. As stated below the larger particles were swollen, with numerous wall punctures. Since the particle interior is accessible to the reacting O_2 at these conditions, the rate of preoxidation becomes independent of particle size.

D. Particle Density

The particle densities of the 28 x 35 mesh product coals from the Hopcalite series were measured in mercury at 1 atm pressure. Results are shown in Figure 36.

High particle density from the preoxidizer is desirable when the coal is gasified, from the kinetic standpoint of maintaining high carbon inventory and gas throughput in the gasifier vessel. From this point of view, the data in Figure 36 show that the preoxidation temperature should be as low as practicable and that the inlet O_2 partial pressure should be as high as practicable. Increasing O_2 partial pressure, at a given temperature, increases the rate of preoxidation and, presumably, that rate at which the particle surface becomes rigid and resistant to swelling.

Microscopic examination of the 28 x 35 mesh, 700-750°F products showed that all the particles, except those rich in fusinoids, were rounded, swollen, and punctured by small holes. The particle interiors had a sponge-like structure. In contrast, none of the 65 x 100 mesh products showed any swelling or wall penetration. The particle densities (Table 40) of the 65 x 100 mesh products were higher than that of the raw coal.

The effects of temperature and particle size upon preoxidized coal density were confirmed by results of the continuous unit which are given in Figure 37 for Ireland Mine coal. Similar trends were observed for Illinois No. 6 coal, but the data scattered badly due to the presence of dense, rocky material in the coal.

* After correction from 730 to 750°F by means of activation energies calculated from Runs 19 and 20.

Figure 36

PARTICLE DENSITY AFTER PREOXIDATION

28 x 35 MESH COAL

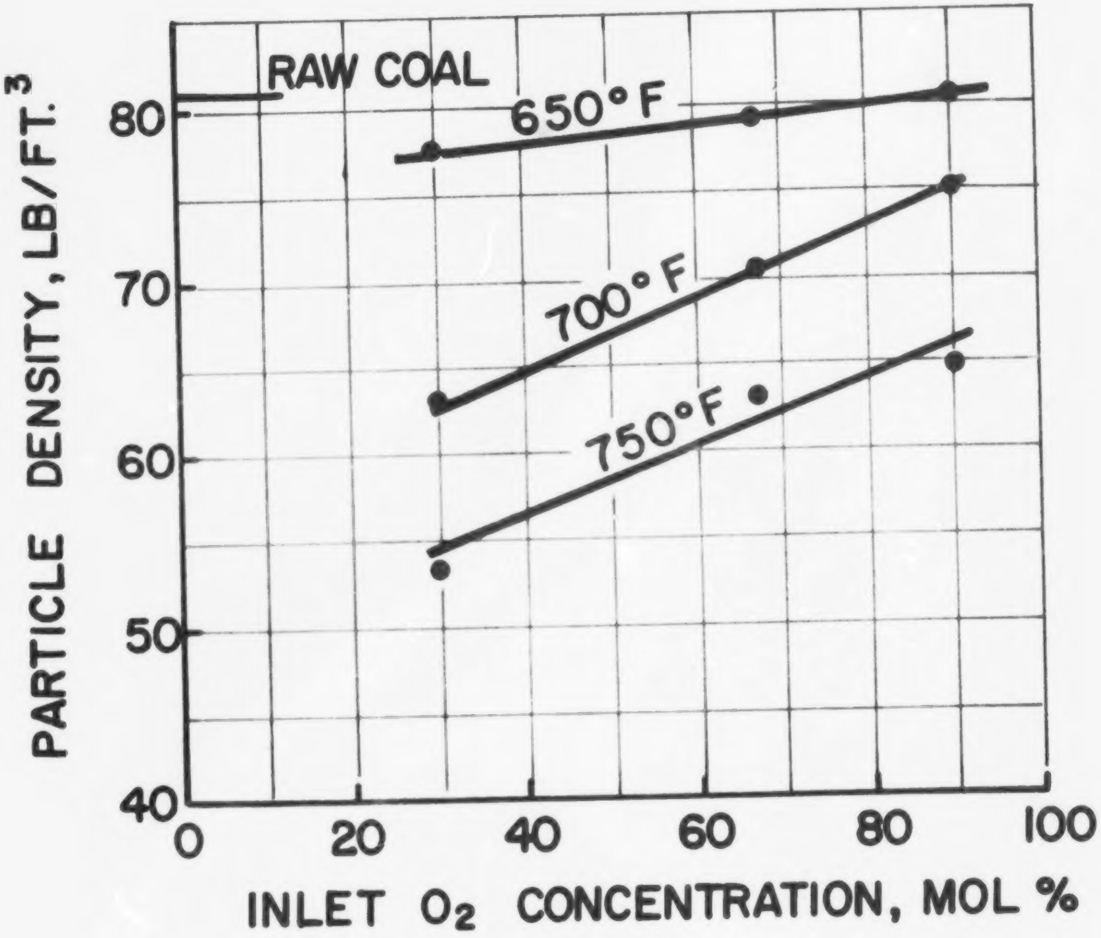
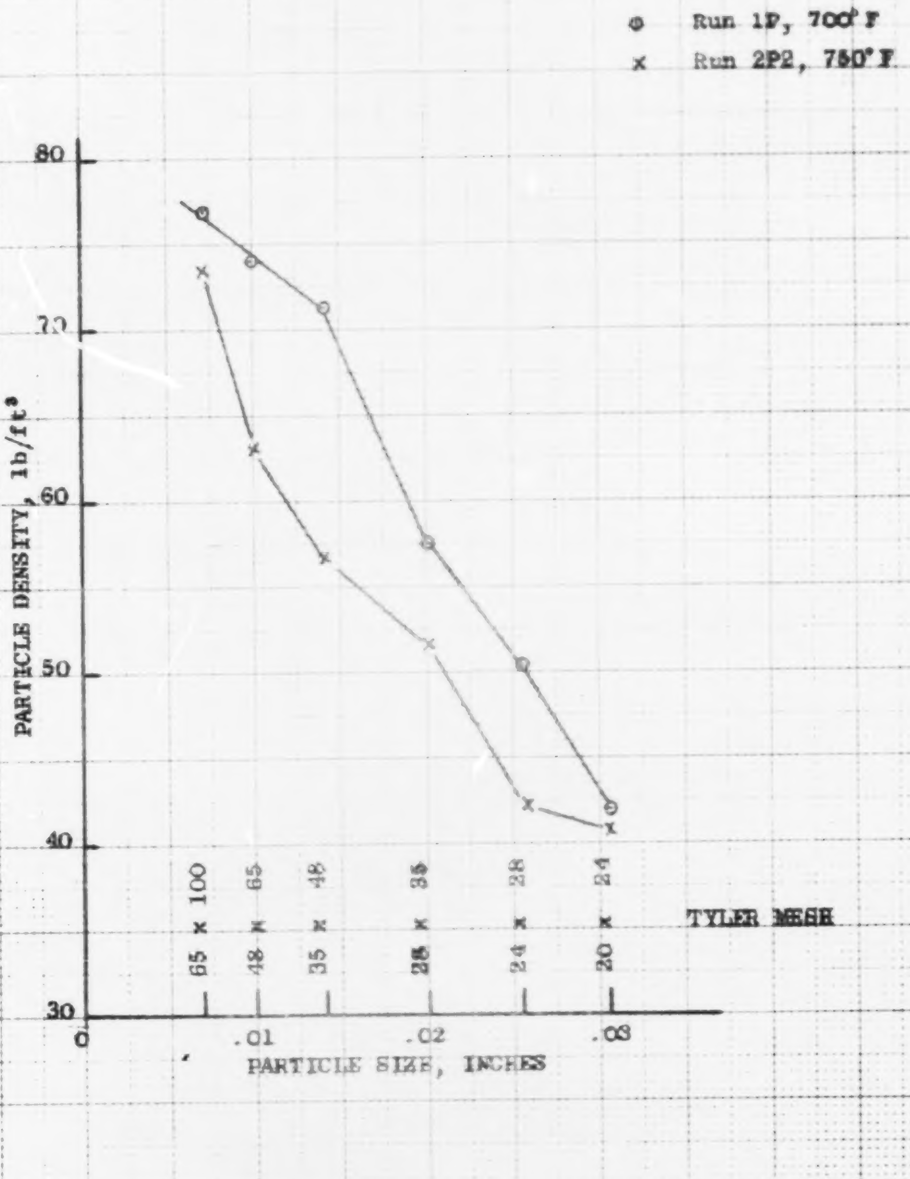


FIGURE 37

PARTICLE DENSITIES FOR PREOXIDIZED IRELAND MINE COAL



E. Dehydrogenation

Decaking of coal by preoxidation involves dehydrogenation. The amount of hydrogen removed from the coal by reaction to form water was calculated for each run from the H_2O absorber weight gain. Results are plotted in Figure 38. The data points are not delineated with respect to run conditions because examination of the individual data points showed no trends. Any separate effects of temperature, O_2 inlet concentration, or particle diameter are confounded, since the higher levels of preoxidation are associated with higher levels of temperature and/or O_2 concentration.

F. Coal Weight Loss

In the temperature range of current interest (600-800°F) for preoxidation, the coal decomposes thermally in an inert atmosphere, with loss of volatile products. Preoxidation inhibits the loss of volatile materials by polymerizing them. At the lower temperatures, enough of the reacting O_2 remains in the polymerized materials to cause a weight gain. In general, the temperatures used in this study were high enough to cause a net weight loss even though some of the reacted O_2 remained in the coal. In Run 20, at 600°F, no net weight change occurred although carbon and hydrogen had been removed from the coal, as shown by the weight gains of the CO_2 and H_2O absorbers.

Percent weight losses for the other runs are plotted in Figure 39. As with the dehydrogenation data, confounding occurred because the higher levels of preoxidation are associated with the higher temperatures. However, there are apparent effects of inlet O_2 concentration and particle diameter on weight loss. As indicated in Figure 39 smaller weight losses occur at increasing O_2 concentrations. Smaller weight losses also occur with the smaller particles. It is not clear whether this behavior is caused by increased retention of the reacted O_2, or by some other mechanism.

FIGURE 38

DEHYDROGENATION VS.
PERCENT PREOXIDATION

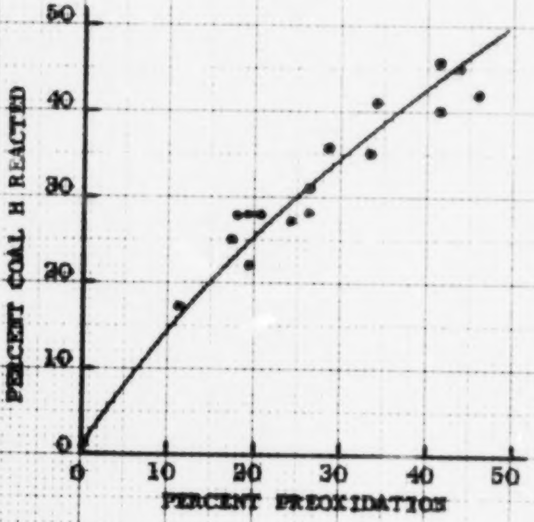
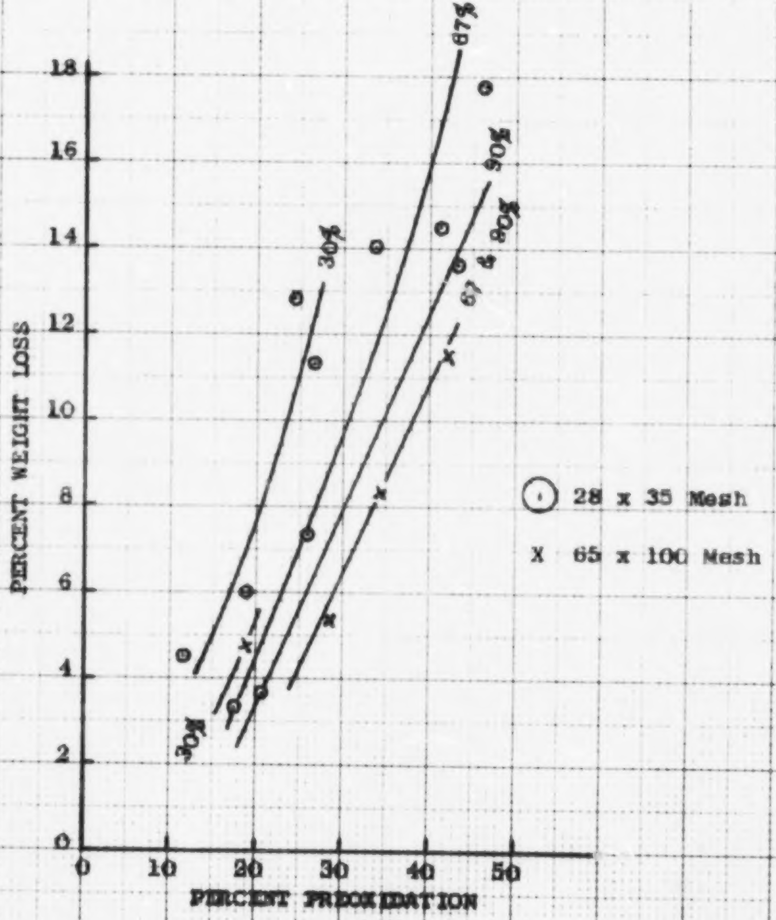


FIGURE 39

COAL WEIGHT LOSS VS.
PERCENT PREOXIDATION



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VIII. APPENDICES

APPENDIX A

Detailed Investment Costs

TABLE A-1

Detailed Plant Investment
Coal Preparation
Case I

Equipment	No. Req'd.	Cost $1000
Coal Storage Bin, 148' x 22' x 16' - 8 Cones C.S.	1	131
Coal Surge Bin, 156' x 12' x 10' - 12 Cones C.S.	1	49
Coal Surge Bins, 18' x 24' x 16' - Double Cone Bottom C.S.	8	160
Compressor, 1050 ACFM, ΔP, 120 psi	1	96
Lock Hoppers, 15' D x 32' C.S.	24	444
Gas Drums, 16' D x 40', 350 psig C.S.	2	80
Conveyors, 36" Belt, 400', 300 tph	2	186
Tripper Conveyors, 100'	4	200
Vibrating Feeders, 83 tph	8	56
Gundlach Crushers, Model 75-2C4R	8	240
Vibrating Feeders, 48 tph	12	84
Gundlach Crushers, Model 50-2C4R	12	324
Collecting Conveyors, 36" Belt, 75'	2	47
Conveyors, 36" Belt, 300', 300 tph	2	150
Distributing Conveyor 440', Multi-tripper points	1	107
Sub-total Major Equipment		2,354
Piling, Foundations, Structural, Electrical, Insulation, Buildings, Piping, and Instrumentation		1,295
Engineering, Supervision, Purchasing, Field Expense, Contractor's Overhead and Profit		1,351
Erected Cost		5,000

TABLE A-2

Detailed Plant Investment
Gasification
Case I

Equipment	No. Req'd.	Cost $1000
Air Heater, 10 MM Btu/hr (incl. air blower)	1	66
Gasifier Steam Superheater - 4,035 ft^2 ea.	6	605
Gasifier Steam W.H.B. - 4,920 ft^2 ea.	6	442
Gasifier Steam W.H.B. - 3,125 ft^2 ea.	2	94
Gasifier Steam BFW Preheater - 2,140 ft^2 ea.	2	64
Gas-Gas Exchanger - 3,880 ft^2 ea.	12	1,070
Preoxidizers, 6'3" OD x 25' OSS C.S.	12	396
Gasifiers, 27'2" OD x 73' OSS C.S. Shell-Refractory-lined	12	6,558
Regenerators, 24'8" OD x 44' OSS C.S. Shell-Refractory-lined	4	1,800
Acceptor Storage Bin, 3000 Tons C.S.	1	82
Acceptor Lock Hoppers, 10'D x 16' C.S.	2	22
Spent Acceptor Lock Hoppers, 7'D x 12' C.S. Shell-Refractory-lined	8	75
Regenerator Ash Lock Hoppers, 7'D x 12' C.S. Shell-Refractory-lined	8	75
Gasifier Internal Cyclones, Refractory-lined	72	1,040
Gasifier External Cyclones, Refractory-lined	36	792
Regenerator External Cyclones, Refractory-lined	8	200
Black Water Pumps, 4250 gpm	2	32
Gasifier Steam Feed Water Pumps, 1463 gpm	2	32
Air Compressors - 68,548 mph ea.	2	8,400
Lift Gas Compressor - 4,528 mph	1	102
Claus Gas Recycle Compressors - 34,790 mph ea.	2	608
Acceptor Air Compressor	1	132
Feed Screws	12	72
Char Rotary Feeders	12	131
Char Rotary Feeders	12	48
Acceptor Rotary Feeders	16	240
Acceptor Rotary Feeders	8	32
Ash-Acceptor Rotary Feeders	4	128
Make-Up Acceptor Rotary Feeders	2	16
Acceptor Feed Belt, 150 tph	1	61
Sub-total Major Equipment		23,415
Piling, Foundations, Structural, Electrical, Insulation, Buildings, Piping, and Instrumentation		15,785
Engineering, Supervision, Purchasing, Field Expense, Contractors Overhead and Profit		15,600
Start-up Steam Plant (Turn-key)		1,200
Erected Cost		56,000

TABLE A-3

Detailed Plant Investment
Sulfur Recovery-Solids Disposal
Case I

Equipment	No. Req'd	Cost $1000
Gas-Gas Exchangers 2,880 ft.^{2} ea.	8	207
Gas-Gas Exchangers 2,610 ft.^{2} ea.	8	188
Sulfur Condensor 6,570 ft.^{2} ea.	16	945
Sulfur Condensor 5,375 ft.^{2} ea.	16	774
Gasifier Steam BFW Preheater 6,200 ft.^{2} ea.	2	74
Reductors 9'3" O.D. x 11'2" OSS C.S.	4	130
First-Stage Claus Converters 13'2" O.D. x 16' OSS C.S.	4	264
Second-Stage Claus Converters 13'2" O.D. x 16' OSS C.S.	4	264
Coalescers	8	128
Acceptor Slurry Stripping Tower 5'6" O.D. x 24' w/Agit.	1	36
Sulfur Storage Tank 10'D x 30'	1	10
Sulfur Melt Tank	1	8
Lock Hoppers	2	22
Sump 5000 gal.	1	6
Water Pumps 3260 GPM	3	18
Sulfur Pumps 80 GPM	2	8
Slurry Pumps 460 GPM	4	30
Hydroclone Overflow Pumps 3960 GPM	2	32
Stripping Gas Compressor	1	102
Electrostatic Precipitators	4	350
Hydroclones 24" D	4	24
Black Water Pond	1	620
Sub-total Major Equipment		4,240
Piling, Foundations, Structural, Electrical, Insulation, Buildings, Piping, and Instrumentation		3,680
Engineering, Supervision, Purchasing, Field Expense, Contractor's Overhead and Profit		3,080
Erected Cost		11,000

TABLE A-4

Detailed Plant Investment
Coal Preparation
Case III

Equipment	No Req'd	Cost $1000
Coal Storage Bin 148' x 22' x 16' - 8 Cones C.S.	1	131
Coal Surge Bin 156' x 12' x 10' - 12 Cones C.S.	1	49
Coal Surge Bins 18' x 24' x 16' Dbl. Cone Bottom C.S.	8	160
Compressor 3230 ACFM △P 150 psi	1	130
Lock Hoppers 18'D x 20' C.S.	16	222
Gas Drums 15'D x 30' 150 psig. C.S.	1	14
Conveyors 36" Belt 400' 300 TPH	2	186
Tripper Conveyors 100'	4	200
Vibrating Feeders 83 TPH	8	56
Gundlach Crushers Model 75-2C4R	8	240
Vibrating Feeders 48 TPH	12	84
Gundlach Crushers Model 50-2C4R	12	324
Collecting Conveyors 36" Belt 75'	2	47
Conveyors 36" Belt 300' 300 TPH	2	150
Distributing Conveyor 440' Multi-tripper points	1	107
Sub-total Major Equipment		2,100
Piling, Foundations, Structural, Electrical, Insulation, Buildings, Piping, and Instrumentation		1,180
Engineering, Supervision, Purchasing, Field Expense, Contractor's Overhead and Profit		1,220
Erected Cost		4,500

TABLE A-5

Detailed Plant Investment
Gasification
Case III

Equipment	No. Req'd	Cost $1000
Air Heater Duty 73 MM Btu/hr. (Incl. Air Blower)	1	298
Gasifier Steam Superheater 3340 ft.^{2} ea.	8	615
Gasifier Steam W.H.B. 3540 ft.^{2} ea.	8	651
Gasifier Steam BFW Preheater 1730 ft.^{2} ea.	2	45
Plant Steam Superheater 2890 ft.^{2} ea.	6	434
Plant Steam W.H.B. 4300 ft.^{2} ea.	6	593
Plant Steam BFW Preheater 3220 ft.^{2} ea.	2	90
Preoxidizers 12' OD x 28' OSS C.S.	16	451
Gasifiers 50' OD x 119' OSS C.S. Shell-Refractory-Lined	16	13,664
Regenerators 35' OD x 38' OSS C.S. Shell-Refractory-Lined	8	2,680
Acceptor Storage Bin 3000 tons C.S.	1	82
Gasifier Internal Cyclones Refractory-Lined	96	1,387
Gasifier External Cyclones " "	96	1,351
Regenerator External Cyclones " "	96	1,412
Condensate Pump 1320 GPM	2	66
Blow-down Pumps 80 GPM	2	12
Gasifier Steam Feed Water Pumps 1230 GPM	2	22
Plant Steam Feed Water Pumps 80 GPM	2	16
Air Compressors 70,498 MPH Steam Turbine Drive 80,000 HP ea.	2	5,120
Lift Gas Compressor 6420 MPH	1	480
Acceptor Air Compressor	1	42
Feed Screws	16	96
Ejectors	16	448
Char Rotary Feeders	16	128
Char Rotary Feeders	16	128
Acceptor Rotary Feeders	32	480
Acceptor Rotary Feeders	16	48
Char Rotary Feeders	8	128
Acceptor Feed Belt 150 TPH	1	61
Acceptor Rotary Feeders	8	24
Sub-total Major Equipment		31,052
Piling, Foundations, Structural, Electrical, Insulation, Buildings, Piping, and Instrumentation		22,748
Engineering, Supervision, Purchasing, Field Expense, Contractors Overhead and Profit		21,500
Start-up Steam Plant (Turn-key)		1,200
Erected Cost		76,500

TABLE A-6

Detailed Plant Investment
Sulfur Recovery - Solids Disposal
Case III

Equipment	No. Req'd	Cost $1000
Gasifier Steam BFW Preheaters 2850 $ft.^2$ ea.	2	57
Plant Steam BFW Preheaters 4125 $ft.^2$ ea.	4	182
Sulfur Condensers 5000 $ft.^2$ ea.	16	720
Water Coolers 200 $ft.^2$ ea.	2	3
Reductors 13'9" OD x 16'6" OSS C.S.	4	142
Claus Convertors 17'3" OD x 20'9" OSS C.S.	4	224
Coalescer 10' OD x 14' OSS C.S.	4	52
Acceptor Slurry Stripping Tower 5'6" OD x 24' w/Agit.	1	36
Sulfur Storage Tank 10' D x 30'	1	10
Sulfur Melt Tank	1	8
Sump 5000 Gal.	1	6
Water Pumps 4520 GPM	3	21
Sulfur Pumps 80 GPM	2	8
Slurry Pumps 470 GPM	4	32
Black Water Pumps 4370 GPM	2	32
Hydroclone Overflow Pumps 4050 GPM	2	32
Stripping Gas Compressors	2	20
Electrostatic Precipitators	8	1,660
Hydroclones 24" D	4	24
Black Water Pond	1	540
Sub-total Major Equipment		3,809
Piling, Foundations, Structural, Electrical, Insulation, Buildings, Piping, and Instrumentation		3,391
Engineering, Supervision, Purchasing, Field Expense, Contractor's Overhead and Profit		2,800
Erected Cost		10,000

APPENDIX B

Discussion of Gasification Design Basis

A. Kinetic Restraints

Design conditions chosen for this feasibility study were noted in Section III-B-1. These conditions were based on information for similar processes which have been accumulated by Consol over a period of 22 years. However, no data are available for the exact operating conditions required in the fuel gas version of the CO_2 Acceptor Process. A major objective of Phases II-A and II-B of the contract is to acquire kinetic data to provide a firm design basis. An outline of the design conditions which are affected by kinetics of the various reactions is shown in Table B-1 for Case I.

B. Effect of Variations in Design Conditions

Since no data are yet available for the exact operating conditions required by the process, a major effort to optimize the gasification conditions would be premature. However, a limited study was done to guide the choice of design conditions which were used in this feasibility study.

In the guideline study, the criterion was the overall heat rate for an integrated power station using the supercharged boiler cycle. The steam cycle conditions were substantially those shown in Table 24. However, for simplicity, no regenerative feed-water heating was employed. All feed-water heating duty was supplied by the expander exhaust gas.

The base case was the same as Case I, described elsewhere in this report, except for the method of feed-water heating. The required heat and material balance calculations are extremely tedious and a computer program, described in Appendix D, was developed to perform the calculations. Results are summarized in Table B-2.

As would be expected, higher inlet gas temperature to the turbine brings a significant reduction in heat rate. Therefore, the 1800°F temperature level was chosen for this study. The impact of the other variables which were considered is relatively small. Detailed design and cost estimates, which are beyond the scope of this work, would be necessary for true optimization with respect to power costs which are influenced by vessel volumes of the gasifier and boiler, by frame sizes for the expanders and compressors, etc.

Another process restriction which probably would reduce the power cost is to preheat the process air by exchange with the hot gas streams. However, this restriction was not considered in the guideline study.

TABLE B-1

Outline of Gasification Design Conditions Affected by Reaction Kinetics - Case I

	Assumed Design Conditions	Process Variable to be Changed to Increase Design Value
Preoxidation	Severity adequate to prevent caking in gasifier	Increase severity by increasing temperature and/or retention time
% preoxidation(1)	8.5	
Temperature, °F	800	
Coal retention time, minutes	10	
Gasification		
Fixed carbon gasification rate, atoms C gasified/atom C in bed/min.	75×10^{-4} at 1700°F	Increase bed temperature and/or inlet steam rate
Methane yield	.042 mols CH_4/atom total carbon gasified(2)	Not independently controllable. Revise correlation on basis of Phase II-B data
Fixed carbon burnoff level	65%	Increase air to gasifier
ΔP_{CO_2} driving force for acceptor reaction, atm.	0.7 at mid-point of bed	Increase system pressure
ΔP_{H_2S} driving force for acceptor reaction	Equilibrium value for $CaO + H_2S = CaS + H_2O$ (96% removal of sulfur)	Decrease steam content of product gas and/or bed temperature
Regeneration		
ΔP_{CO_2} driving force for calcining acceptor	1.0 atm in exit gas	Increase bed temperature
ΔP_{SO_2} driving force for sulfur rejection	30% of equilibrium SO_2 pressure for $3/4\ CaSO_4 + 1/4\ CaS = CaO + SO_2$	Increase CO partial pressure by decreasing air/(fuel carbon) ratio

(1) 100 (lb O_2 consumed)/(lb MF coal).

(2) Based on correlation of data in work described in reference 4.

TABLE B-2

Effect of Process Variables on Overall Station Heat Rate
Case I with Supercharged Boiler Cycle

<table>
<tr><th>Case</th><th>Base</th><th>1</th><th>2</th><th>3</th><th>4</th><th>5</th><th>6</th><th>7</th><th>8</th><th>9</th><th>10</th></tr>
<tr><td>% Fixed Carbon Burnoff</td><td>65</td><td>51</td><td colspan="9">← 65 →</td></tr>
<tr><td>Fixed Carbon Gasification Rate, atoms C gasified/atm C in bed/min x 10^4</td><td>75</td><td>75</td><td>102</td><td>56</td><td>75</td><td>102</td><td colspan="5">← 75 →</td></tr>
<tr><td>Intercooling on Total Station air</td><td colspan="4">← Yes →</td><td colspan="3">← No →</td><td colspan="4">← Yes →</td></tr>
<tr><td>% Excess Air to Boiler</td><td colspan="10">← 5 →</td><td>50</td></tr>
<tr><td>System Pressure, Atmospheres</td><td colspan="8">← 15 →</td><td>16</td><td>15</td><td>15</td></tr>
<tr><td>Expander Inlet Temperature, °F</td><td colspan="6">← 1800 →</td><td>1600</td><td>1600</td><td colspan="3">← 1800 →</td></tr>
<tr><td>Acceptor Make-up Rate, %</td><td colspan="9">← 0.5 →</td><td>1.0</td><td>0.5</td></tr>
<tr><td>Overall Heat Rate, % Difference from Base Case</td><td>--</td><td>+.71</td><td>-.41</td><td>+.38</td><td>-.28</td><td>-.56</td><td>+2.42</td><td>+3.00</td><td>+.31</td><td>+.13</td><td>-1.54</td></tr>
</table>

APPENDIX C

Data and Procedure Used for Sulfur Removal and Recovery

A. Introduction

Presented here is a body of self-consistent thermodynamic data for use in equilibrium and heat balance calculations.

When catalyzed by an active alumina, the reductor and Claus reactions approach equilibrium closely. At the higher temperatures (1600-1950°F) involved in the sulfur-removal reactions, previous operations of the continuous CO_2 Acceptor Gasification Unit have shown, qualitatively, that equilibrium also is approached closely. Apparently, catalysis is by the coal ash and/or the sulfur acceptor. Thus, equilibrium calculations provide a reliable guide to actual plant performance.

A computer program was developed to perform the extremely tedious equilibrium calculations over the temperature range of 250 to 2000°F.

B. Method for Calculation of Equilibrium in the Reductor and Claus Reactors

A system containing reactants and products for a specified temperature and total pressure can be defined in several ways. The one chosen for this work is described below.

Equilibrium is assumed for the following reactions:

$$CO + 1/2\ S_2 = COS \qquad (1)$$

$$H_2 + 1/2\ S_2 = H_2S \qquad (2)$$

$$2\ H_2 + SO_2 = 1/2\ S_2 + 2\ H_2O \qquad (3)$$

$$2\ CO + SO_2 = 1/2\ S_2 + 2\ CO_2 \qquad (4)$$

$$3\ S_8 = 4\ S_6 \qquad (6)$$

$$S_8 = 4\ S_2 \qquad (8)$$

These six reactions, in conjunction with the four elemental balances listed below, are sufficient to define the system.

An eleventh possible component, CS_2, has been neglected in this work because preliminary calculations showed that, at the partial pressures of the reactants and products involved in CS_2 production in the processes considered here, a negligible amount of CS_2 can be formed in the sulfur-removal portion of the process (1400 to 2000°F). Furthermore, at reductor and Claus reactor conditions (1000 to 250°F), the kinetics of the reaction $2\ COS = CO_2 + CS_2$ are very poor.(13)

$$[S] = 2\ S_2 + 6\ S_6 + 8\ S_8 + SO_2 + H_2S + COS \tag{9}$$

$$[C] = CO_2 + CO + COS \tag{10}$$

$$[O] = 2\ CO_2 + 2\ SO_2 + CO + H_2O + COS \tag{11}$$

$$[H] = H_2O + H_2 + H_2S \tag{12}$$

where: The chemical symbol for a substance equals the number of mols of that substance in the product gas,

[S] = atoms total sulfur in the feed gas.
[C] = atoms total carbon in the feed gas.
[O] = atoms total oxygen in the feed gas.
[H] = mols total H_2 in the feed gas.

Equilibria in the six reactions can be expressed as follows, where the various K's are equilibrium constants obtained from experimental data or calculated from modern free energy values, and P is the system pressure, atm:

$$K_1 = \frac{COS}{CO\ (S_2)^{1/2}} \left(\frac{W}{P}\right)^{1/2} \tag{13}$$

$$K_2 = \frac{H_2S}{H_2\ (S_2)^{1/2}} \left(\frac{W}{P}\right)^{1/2} \tag{14}$$

$$K_3 = \left(\frac{H_2O}{H_2}\right)^2 \frac{(S_2)^{1/2}}{SO_2} \left(\frac{W}{P}\right)^{1/2} \tag{15}$$

$$K_4 = \left(\frac{CO_2}{CO}\right)^2 \frac{(S_2)^{1/2}}{SO_2} \left(\frac{W}{P}\right)^{1/2} \tag{16}$$

$$K_6 = \frac{(S_6)^4}{(S_8)^3} \left(\frac{P}{W}\right) \tag{17}$$

$$K_8 = \frac{(S_2)^4}{S_8} \left(\frac{P}{W}\right)^3 \tag{18}$$

In the above equations, W equals the total mols of product gas, which can be expressed as:

$$W = [C] + [H] + I + S_8 + S_6 + S_2 + SO_2. \tag{19}$$

where: I = mols inert gas (usually N_2).

Solution of equations (9) through (19) is accomplished as follows:

1. An initial value for S_2 is assumed, using the data in Table C-I as a guide. The system is not sensitive to small changes in the value of W and the initial value is obtained by substituting the term [S]/3, for the terms, $S_8 + S_6 + S_2 + SO_2$, in equation (19).

2. Equations (10) through (18) are solved algebraically and the sulfur balance {equation (9)} is determined.

3. The value for S_2 is then changed in proportion to the magnitude and direction of deviation from 100% closure of the sulfur balance. W is calculated from the values of $S_8 + S_6 + S_2 + SO_2$ just obtained in Step 2.

4. Steps 2 and 3 are repeated until the sulfur balance closes to within 100.00 ± .05%.

For multi-stage Claus reactor calculations, a trial and error method involving the reactor temperature is required. For the first stage, a reactor temperature is assumed and equations (9) through (19) are solved as described above. The partial pressure of elemental sulfur then is calculated and compared with the vapor pressure of liquid sulfur. The temperature then is changed and the calculation is repeated until the sulfur partial pressure is about 20% less than the vapor pressure. This procedure was adopted in order to allow for errors in estimation of the heat exchange requirements for the feed gas. The Claus reaction is moderately exothermic. Also, condensation of liquid sulfur on the catalyst destroys its activity. Between stages, the product gas is cooled to condense substantially all the elemental sulfur. For the second and third (if needed) stage calculations, the elemental sulfur formed in the preceding stage is subtracted from the value of [S] in equation (9) and the entire process is repeated, except that the final temperature corresponds to a sulfur partial pressure of about 10% less than the vapor pressure. The closer approach to saturation temperature was used because the heat of reaction is much lower in the second and third stages. The complete calculation for a three-stage process requires, typically, about 200 iterations of Step 2, above.

C. Discussion

As noted in the description of the sulfur recovery section for Case I (Section III-B-2), the kinetics of reactions (1) through (4) become very slow at Claus reactor temperatures (<600°F). The actual Claus reactions, which kinetically are rapid when catalyzed by alumina, can be expressed for diatomic sulfur as:

$$2\,H_2S + SO_2 \;=\; 3/2\,S_2 + 2\,H_2O,$$

and

$$2\,COS + SO_2 \;=\; 3/2\,S_2 + 2\,CO_2.$$

Equilibrium constants for the above Claus reactions are given by $K_3/(K_2)^2$ and $K_4/(K_1)^2$, respectively. Thus, a system at equilibrium with respect to reactions (1) through (4) also is at equilibrium with respect to the Claus reactions.

Case I is unique in application of the computer program to the sulfur recovery section. In this case CO and H_2 are brought into the Claus reactors with that portion of the gas which bypasses the reductor. This CO and H_2 is treated as an inert gas in the calculations.

D. Sources of Data, Gas Reactions

Nearly all of the free energy values, heat capacities at zero pressure, and heats of formation were taken from the JANAF tables.(14) The equilibrium constants were calculated from the free energy values. Exceptions were:

1. The equilibrium constants for reactions (6) and (8) were taken from the experimental data of Preuner and Schupp.(15)

2. The heat capacity and heat of formation for S_6 were calculated from the JANAF data for S_8 and the equilibrium constants measured by Preuner and Schupp for reaction (6).

The vapor pressure and heat of vaporization of liquid sulfur were taken from West(16) and the heat content of steam above liquid water were values given in the Keenan and Keyes steam tables.

E. Presentation of Data, Gas Reactions

The effects of temperature and sulfur partial pressure are shown in Table C-1. The first entries for each temperature through 600° F correspond to the vapor pressure of liquid sulfur. The data were calculated from equations (17) and (18), taking the sum of the mol fractions of S_8, S_6, and S_2 as unity.

Equilibrium constants are shown in Table C-2 and numerical values of the constants as a function of temperature are in Table C-3. Heat capacities are in Table C-4 and the mean heat capacities above 60° F, derived from the Table C-4 data, are in Table C-5. The heat content of steam above liquid water at 60° F also is given in Table C-5.

Heats of formation and heats of reaction at 25° F are given in Table C-6. Equations for the vapor pressure of liquid sulfur are shown in Figure C-1, which is a plot of the equations. An equation for the heat of vaporization of liquid sulfur also is given in Table C-6.

F. Data for Solids Reactions

Equilibria for the pertinent solids reactions involving calcium compounds are given in Table C-7. Mean heat capacities above 60° F are shown in Table C-8. Heats of formation and heats of reaction are shown in Table C-9.

Sources of these data primarily were experimental measurements of the thermodynamic properties of $CaCO_3$, CaO, CaS and $CaSO_4$ which were supported privately by Consolidation Coal Company during the 1960's.

G. Calculations for Sulfur Rejection in the Regenerator - Case I

Equilibrium for some pertinent sulfur reactions involving lime are plotted in Figures C-2 and C-3.

To calculate the exit gas composition in the regenerator, equilibrium is assumed for the following reactions:

$$2\ CO + SO_2 = 1/2\ S_2 + 2\ CO_2$$

$$CaS + H_2O = CaO + H_2S$$

$$CaS + CO_2 = CaO + COS$$

$$H_2O + CO = H_2 + CO_2$$

These four reactions, in conjunction with the four elemental balances for C, O, H, and S are sufficient to define the system.*

In a manner analogous to the formulation of equations (9) through (19), a system of nine simultaneous equations was set up. Solution of the equations is accomplished as follows:

1. An initial value for the mol ratio CO_2/CO, is assumed.
2. The equations are solved algebraically and the sulfur balance is determined.
3. The value for the CO_2/CO ratio is changed in proportion to the magnitude and direction of deviation from 100% closure of the sulfur balance.
4. Steps 2 and 3 are repeated until the sulfur balance closes to within 100.0 ± 0.1%

The above procedure is incorporated as part of the computer program for the process heat and material balances described in Appendix D.

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* Note that equilibrium is not assumed for the entire system. Rather, the assumption is made that the kinetics of the above reactions are rapid compared with those of back-reactions such as:

$$CaO + 3/4\ S_2 = CaS + 1/2\ SO_2.$$

$$CaSO_4 + S_2 = CaS + 2\ SO_2.$$

TABLE C-I

Effect of Temperature and Sulfur Partial Pressure on Distribution of Sulfur Species

Mol Fraction S_8	Mol Fraction S_6	Mol Fraction S_2	Partial Pressure of Total Sulfur, atm	Atoms Sulfur/Mol Total Sulfur
			250° F	
.8575	.1424	.0001	4.85×10^{-5}	7.714
.8472	.1527	.0001	3.54×10^{-5}	7.684
.8080	.1917	.0003	1.24×10^{-5}	7.615
.7567	.2426	.0007	$.396 \times 10^{-5}$	7.511
.7022	.2963	.0015	$.142 \times 10^{-5}$	7.398
			300° F	
.8249	.1749	.0002	2.33×10^{-4}	7.649
.8000	.1997	.0003	1.25×10^{-4}	7.599
.7473	.2520	.0007	$.401 \times 10^{-4}$	7.492
.6916	.3068	.0016	$.144 \times 10^{-4}$	7.377
.6230	.3735	.0035	$.0482 \times 10^{-4}$	7.232
			400° F	
.7566	.2428	.0006	3.11×10^{-3}	7.511
.7158	.2831	.0011	1.40×10^{-3}	7.427
.6501	.3474	.0025	$.461 \times 10^{-3}$	7.290
.5841	.4108	.0051	$.171 \times 10^{-3}$	7.148
.5072	.4819	.0109	$.0592 \times 10^{-3}$	6.971
			500° F	
.6903	.3080	.0017	2.42×10^{-2}	7.374
.5903	.4045	.0052	$.508 \times 10^{-2}$	7.160
.5204	.4693	.0103	$.192 \times 10^{-2}$	6.999
.4410	.5374	.0216	$.0680 \times 10^{-2}$	6.795
.3682	.5906	.0412	$.0272 \times 10^{-2}$	6.572
			600° F	
.6079	.3873	.0048	9.81×10^{-2}	7.196
.5651	.4274	.0075	5.31×10^{-2}	7.100
.4937	.4913	.0150	2.02×10^{-2}	6.927
.4133	.5558	.0309	$.726 \times 10^{-2}$	6.703
.3401	.6019	.0580	$.294 \times 10^{-2}$	6.448
.2623	.6273	.1104	$.114 \times 10^{-2}$	6.083
.1958	.6163	.1879	$.0511 \times 10^{-2}$	5.640
			700° F	
.4614	.5159	.0227	.1300	6.832
.4146	.5512	.0342	.0724	6.692
.3404	.5955	.0641	.0294	6.424
.2612	.6175	.1213	.0115	6.037
.1934	.6018	.2048	.00517	5.568
.1279	.5380	.3341	.00234	4.919
.0801	.4431	.4768	.00124	4.253
			800° F	
.3287	.5949	.0764	.1825	6.352
.2822	.6075	.1103	.1063	6.123
.2117	.5997	.1886	.0472	5.669
.1423	.5448	.3129	.0211	5.033
.0910	.4582	.4508	.0110	4.378
.0499	.3398	.6103	.00601	3.658
.0264	.2368	.7368	.00378	3.106
			900° F	
.2170	.5970	.1860	.2764	5.690
.1751	.5726	.2523	.1714	5.341
.1164	.5012	.3824	.0859	4.703
.0670	.3899	.5431	.0448	3.962
.0369	.2822	.6809	.0271	3.350
.0176	.1817	.8007	.0171	2.832
.0084	.1149	.8767	.0118	2.510
.0016	.0398	.9586	.00609	2.169
			1000° F	
.0977	.4625	.4398	.3071	4.436
.0572	.3561	.5867	.1749	3.768
.0288	.2423	.7289	.1042	3.142
.0143	.1588	.8269	.0698	2.721
.0063	.0946	.8991	.0475	2.416
.0029	.0571	.9400	.0345	2.246
.0005	.0189	.9806	.0186	2.078
			1200° F	
.0055	.0870	.9075	.546	2.381
.0025	.0523	.9452	.398	2.224
.0010	.0294	.9696	.287	2.124
.0004	.0172	.9824	.216	2.072
.0001	.0055	.9944	.120	2.022
.0000	.0007	.9993	.0328	2.003

TABLE C-2

Equilibrium Constants for Gas Reactions

Temperature in °Kelvin

$$\ln K = A + BT + CT^2 + D/T$$

		A	B	$C \times 10^6$	D	Temperature Range, °K
K1	$CO + 1/2\ S_2 = COS$	-15.5992	.0122924	-8.01042	12,017.2	300-900
		-10.3503	.0006332	-.10012	11,327.0	900-1400
K2	$H_2 + 1/2\ S_2 = H_2S$	-3.3749	-.0027184	1.01362	10,039.6	300-900
		-4.9853	-.0007315	.18766	10,480.5	900-1400
K3	$2H_2 + SO_2 = 1/2\ S_2 + 2H_2O$	-4.8008	.0035970	-2.26137	15,353.9	300-900
		-1.4871	-.0021610	.43958	14,786.0	900-1400
K4	$2CO + SO_2 = 1/2\ S_2 + 2CO_2$	-15.7562	.0052624	-2.65427	25,277.0	300-900
		-14.3343	.0015907	-.35856	25,314.0	900-1400
K6	$3S_8 = 4S_6$	21.9713	-.003582	3.5055	-15,135.4	300-800
K8	$S_8 = 4S_2$	-33.2782	.132324	-63.9707	-29,596.0	300-800
K5	$2H_2S + SO_2 = 3/2\ S_2 + 2H_2O$	1.99482	.0089502	-4.6424	-4370.84	300-900

TABLE C-3

Numerical Values of Equilibrium Constants for Table C-2

Temperature in °F

°F	K1	K2	K3	K4	K6	K8	K5
200	1.05×10^{8}	1.55×10^{9}	1.80×10^{16}	5.27×10^{21}	3.17×10^{-8}	2.03×10^{-29}	.000749
300	1.66×10^{7}	2.76×10^{8}	1.45×10^{14}	8.45×10^{19}	3.88×10^{-7}	2.61×10^{-26}	.00190
400	8.04×10^{5}	1.57×10^{7}	2.26×10^{12}	9.35×10^{16}	2.44×10^{-5}	5.75×10^{-21}	.00914
500	73,900	1.60×10^{6}	8.36×10^{10}	4.35×10^{14}	6.62×10^{-4}	1.56×10^{-16}	.0325
600	10,600	2.49×10^{5}	5.72×10^{9}	5.62×10^{12}	9.81×10^{-3}	8.48×10^{-13}	.0925
700	2,090	5.27×10^{4}	6.15×10^{8}	1.55×10^{11}	.0937	1.26×10^{-9}	.221
800	517	1.42×10^{4}	9.29×10^{7}	7.55×10^{9}	.644	6.30×10^{-7}	.461
900	152	4,620	1.82×10^{7}	5.73×10^{8}	3.43	1.25×10^{-4}	.856
1000	50.3	1,750	4.39×10^{6}	6.16×10^{7}	15.1	.0108	1.44
1200	7.06	353	4.01×10^{5}	1.55×10^{6}	188	9.54	3.22
1400	3.18	99.6	6.35×10^{4}	9.15×10^{4}	--		6.39
1600	1.15	35.9	1.38×10^{4}	9,280	--		10.7
1800	.501	15.5	3,900	1,420	--		16.3
2000	.251	7.64	1,340	296	--		22.9

TABLE C-4

Heat Capacities at Zero Pressure

Temperature in °Kelvin

$$C_p = A + BT + CT^2 + DT^3 + E/T^2$$

	A	B	C x 10^6	D x 10^9	E	Temp. Range, °K
S_8	28.51	.03771	-25.25			300-800
S_6	23.13	.01852	-3.137			300-1000
S_2	8.633	.000272			-87,826	300-1400
COS	9.678	.005978	-.1903		-123,658	
H_2S	7.344	.001851	3.364	-1.623		
SO_2	5.855	.01534	-11.002	2.842		
N_2	7.098	-.001431	3.490	-1.348		
CO	5.551	.003185	-.8609		48,814	
CO_2	5.101	.015568	-10.236	2.552		
H_2	7.219	-.000674	.6840		-16,003	
$H_2O_{(g)}$	7.757	.0000003	3.219	-1.125		

TABLE C-5

Mean Heat Capacities Above 60°F, Gases

Temperatures in °F

$$\bar{C}p = A + BT + CT^2 + DT^3$$

Btu/Lb Mol/°F

	A	B	C x 10^6	D x 10^9	Temp. Range, °F
S_8	36.903	.006734	-2.598		60-1000
S_6	27.940	.004678	-.321		60-1300
S_2	7.680	.001251	-.640	.126	60-2000
COS	9.718	.003033	-1.096	.177	
H_2S	8.065	.000916	.214	-.070	
SO_2	9.274	.002798	-.898	.121	
N_2	6.941	.000037	.252	-.058	
CO	6.925	.000105	.266	-.070	
CO_2	8.632	.002958	-.844	.109	
H_2	6.910	.000135	-.044	.022	
H_2O (g)	7.974	.000410	.239	-.048	↓

Heat content of steam (1 a m) above liquid water at 60°F.

Btu/Lb = $1021.08 + .48540T - 35.83 \times 10^{-6} T^2 + 34.46 \times 10^{-9} T^3$	220-1000
Btu/Lb = $1039.25 + .41916T + 46.60 \times 10^{-6} T^2$	1000-1600 *

* Can be extrapolated accurately to 2000°F.

TABLE C-6

Heats of Formation at 25° C, Gases

cal/g mol

	ΔHf
$S_8(g)$	+24,200
$S_6(g)$	+25,580
$S_2(g)$	+30,840
COS	-33,080
H_2S	-4,880
SO_2	-70,960
CO	-26,416
CO_2	-94,052
$H_2O_{(g)}$	-57,798
$H_2O_{(l)}$	-68,430

Heats of Reaction at 25° C, Gas Reactions

cal/g mol

Reaction		
1	$CO + 1/2\ S_2 = COS$	-22,080
2	$H_2 + 1/2\ S_2 = H_2S$	-20,300
3	$2H_2 + SO_2 = 1/2\ S_2 + 2H_2O_{(g)}$	-29,220
4	$2CO + SO_2 = 1/2\ S_2 + 2CO_2$	-48,890
5	$2H_2S + SO_2 = 3/2\ S_2 + 2H_2O_{(g)}$	+11,380
6	$3S_8 = 4S_6$	+29,720
8	$S_8 = 4S_2$	+99,160
6a	$1/6\ S_6 = 1/2\ S_2$	+11,160
8a	$1/8\ S_8 = 1/2\ S_2$	+12,400
5a	$2H_2S + SO_2 = 1/2\ S_6 + 2H_2O_{(g)}$	-22,090
5b	$2H_2S + SO_2 = 3/8\ S_8 + 2H_2O_{(g)}$	-25,800

Heat of Vaporization of Sulfur

Temperature in °F

$$\text{Btu/Lb} = 211.1 - .3302T + 434.6 \times 10^{-6}\ T^2 - 214.4 \times 10^{-9}\ T^3$$

TABLE C-7

Equilibrium Constants for Solids Reactions

Temperature in °Farenheit

$\ln K = A + BT + CT^3 + D/T$

	A	B	C x 10^6	D	Temperature Range, °F
3/4 $CaSO_4$ + 1/4 CaS = CaO + SO_2	71.5076	-.031250	5.79655	-67,104	1500-2000
1/4 $CaSO_4$ + CO = 1/4 CaS + CO_2	3.0040	-.000751	.01932	4,232	1500-2000
CaO + H_2S = CaS + H_2O	17.5646	-.0092926	1.7716	-1,070	1300-2000
CaO + 3/4 S_2 = CaS + 1/2 SO_2	-7.22273	.0025330	-.36880	15,200	1500-2000
$CaSO_4$ + S_2 = CaS + 2 SO_2	-6.55015	.0118713	-1.86265	-12,576	1500-2000
$CaCO_3$ = CaO + CO_2	-.840165	.0074015	-1.28243	-12,932	all
$CaCO_3$ + H_2S = CaS + CO_2 + H_2O	16.6700	-.0018535	.480912	-13,977.2	1100-2000

TABLE C-8

Mean Heat Capacities Above 60°F, Solids

Temperature in °Farenheit

$$\bar{C}p = A + BT + CT^2$$

Btu/lb Mol/°F

	A	B	C x 10^6	Temperature Range, °F
$CaSO_4$	23.57	.00645	-.0383	60-2000
CaS	10.14	.00230	-.502	
CaO	10.98	.00071	-.0772	
MgO	9.77	.00081	-.0387	
$MgO \cdot CaO$	20.75	.00152	-.116	
$MgO \cdot CaCO_3$	30.64	.00506	-.193	↓
Char Carbon	1.447	.00291	-.643	
		Btu/lb/°F		
Ash*	.207	.000029	-.0034	60-2000
		Heat Content Above 60°F Btu/lb		
Coal**	-12.2	.1879	.00025	60-800

* Also used for impurity content of acceptor.

** Including products of pyrolysis.

TABLE C-9

Heats of Formation at 25° C, Solids

cal/g mol

	ΔH_f
$CaSO_4$	-344,090
CaS	-113,550
$CaCO_3$	-228,280
CaO	-151,900

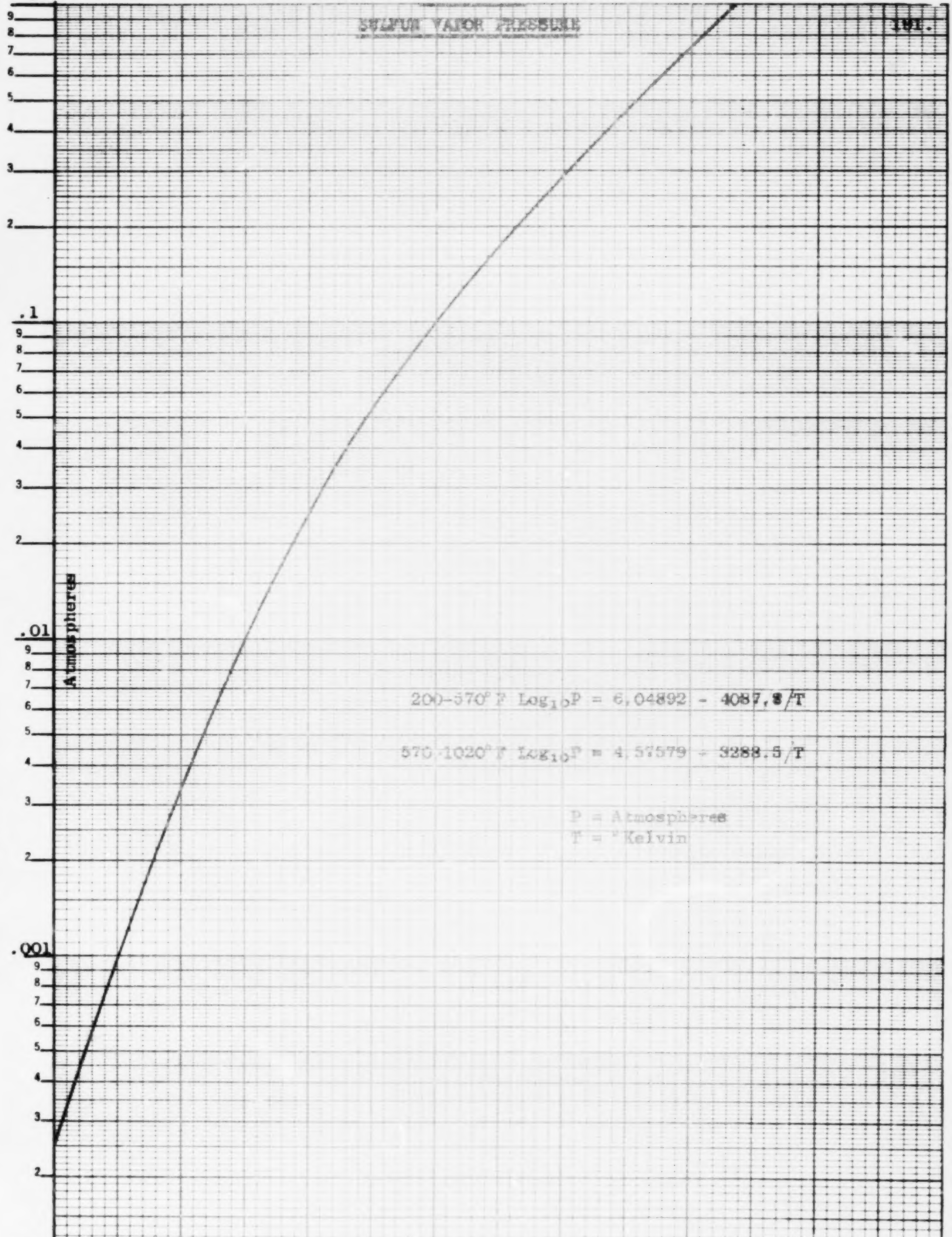
Heats of Reaction at 25° C, Solids Reactions

cal/g mol

$3/4\ CaSO_4 + 1/4\ CaS = CaO + SO_2$	+63,600
$1/4\ CaSO_4 + CO = 1/4\ CaS + CO_2$	-10,002
$CaO + H_2S = CaS + H_2O_{(g)}$	-14,570
$CaO + 3/4\ S_2 = CaS + 1/2\ SO_2$	-20,260
$CaSO_4 + S_2 = CaS + 2\ SO_2$	+57,780
$CaCO_3 = CaO + CO_2$	+42,330
$CaCO_3 + H_2S = CaS + CO_2 + H_2O_{(g)}$	+27,760
$CaS + 3/2\ O_2 = CaO + SO_2$	-109,310
$CaS + 2\ O_2 = CaSO_4$	-230,540

FIGURE C-1

SULFUR VAPOR PRESSURE



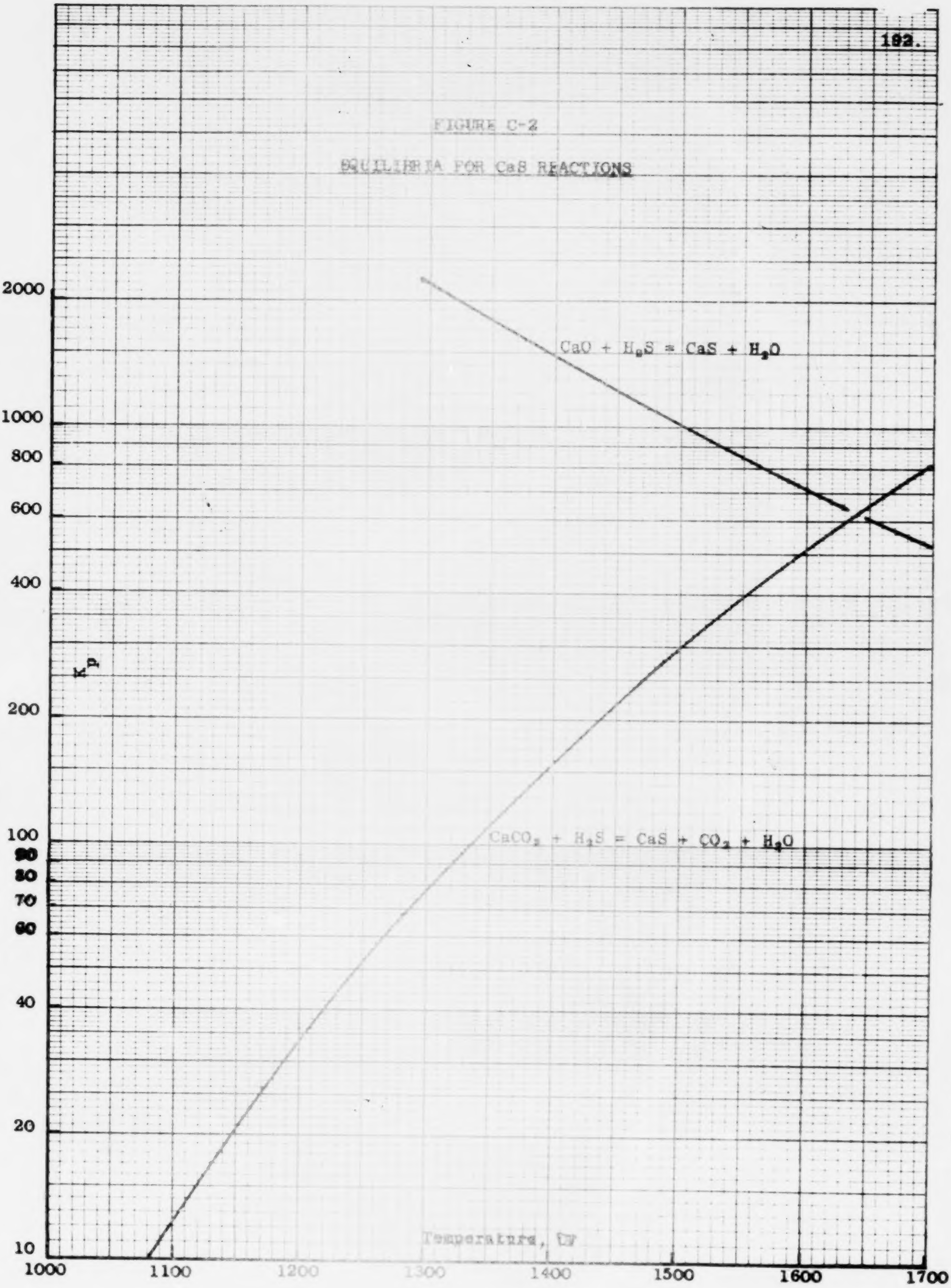
200-570° F $\mathrm{Log}_{10}P = 6{,}04892 - 4087{,}8/T$

570-1020° F $\mathrm{Log}_{10}P = 4.57579 - 3288.5/T$

P = Atmospheres
T = ° Kelvin

FIGURE C-2

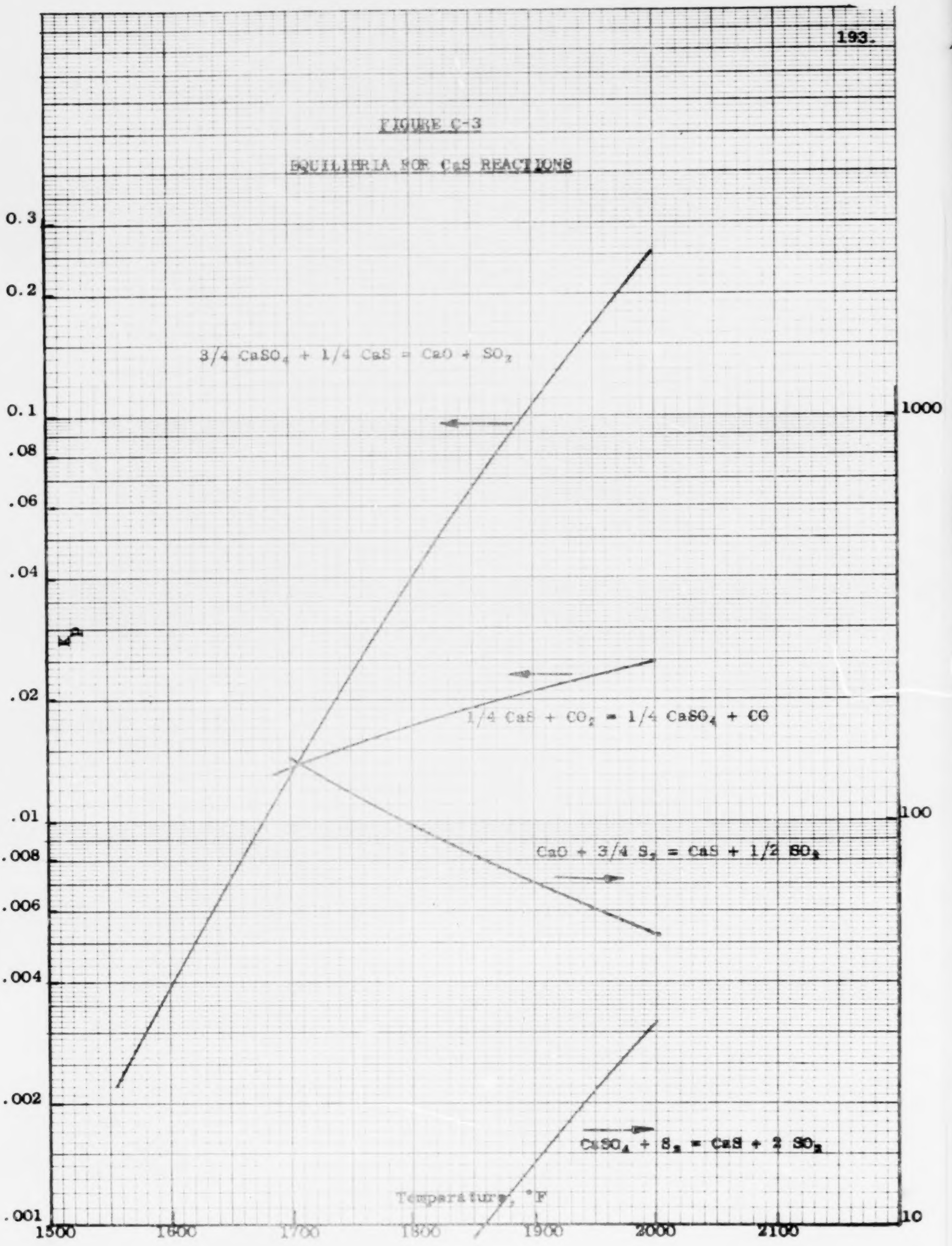
EQUILIBRIA FOR CaS REACTIONS



K&E SEMI-LOGARITHMIC 46 5493
3 CYCLES X 70 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

FIGURE C-3

EQUILIBRIA FOR CaS REACTIONS



K&E SEMI-LOGARITHMIC 46 5493 3 CYCLES X 70 DIVISIONS KEUFFEL & ESSER CO.

APPENDIX D

Description of Computer Program for Process Heat and Material Balances

The computer program described herein was developed to be consistent with the design constraints as set forth on pages 21 and 22.

The elemental balance relationships for the preoxidizer-gasifier system lead to a set of simultaneous equations which cannot be solved analytically. For a given set of process conditions,* the equations are solved by assuming initial values for the following ratios:

$CaCO_3/\Sigma C$ = (mols $CaCO_3$ formed)/(atoms total carbon gasified).
$C/CaCO_3$ = (lb fuel carbon fed to the regenerator)/(mol $CaCO_3$ formed).
$CH_4/\Sigma C$ = (mols CH_4 formed)/(atoms total carbon gasified).

Subsequent steps are shown in Figure D-1, which is a flow diagram for the computer program. After convergence of the preoxidizer-gasifier and regenerator heat balances, and after establishment of the desired partial pressure driving forces for calcining of $CaCO_3$ and for sulfur rejection in the regenerator, the following kinetic restraints are calculated and printed:

1. % burnoff of fixed carbon in the gasifier.
2. ΔP_{CO_2} driving force for recarbonation at the midpoint of the gasifier.
3. Mean gasification rate of fixed carbon.

If any of the kinetic restraints are not at the value chosen for the design basis, new values for specific input variables, as shown below, are selected and the entire calculation is repeated.

	Restraint	Variable Changed
1.	% burnoff	input air to gasifier
2.	ΔP_{CO_2}	system pressure
3.	gasification rate	input steam flow and/or gasifier temperature.

In principle, additional loops could have been incorporated in the program to force convergence on any desired values of the kinetic restraints. However, the additional commands involved would exceed the allowable core capacity of the particular time-sharing computer used by Consol.

After the kinetic restraints are satisfied, all data needed for presentation of the mass and heat balances, such as those in Tables V and VI, are printed.

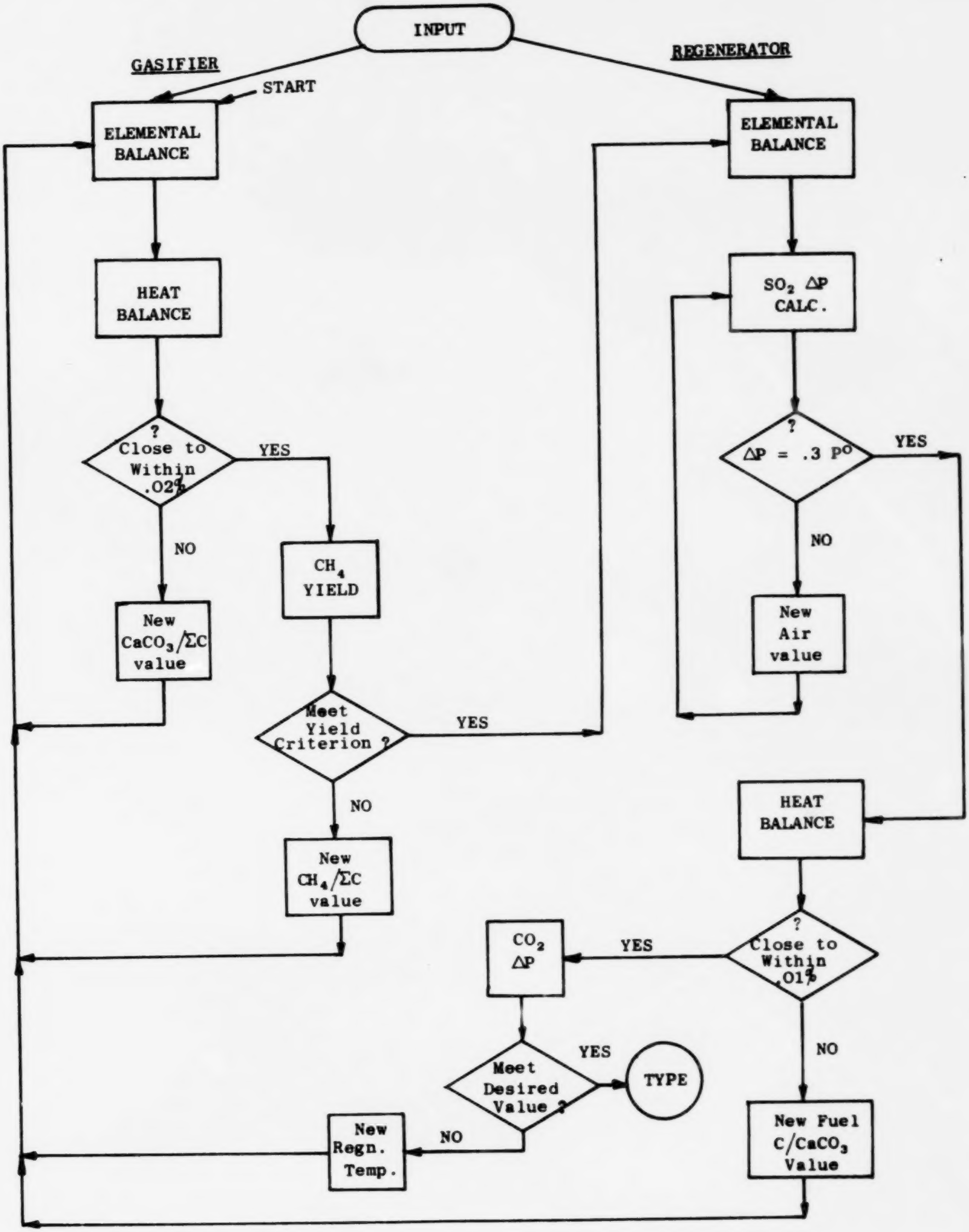
A listing of the program is attached.

---

* Necessary input data involving the process conditions are shown on the first page of the listing of the program. Input data are based on 100 pounds dry coal fed to the preoxidizer.

FIGURE D-1

FLOW DIAGRAM FOR COMPUTER PROGRAM


INPUT
GASIFIER
START
REGENERATOR
ELEMENTAL BALANCE
HEAT BALANCE
? Close to Within .02%
YES
NO
New CaCO$_3$/ΣC value
CH$_4$ YIELD
Meet Yield Criterion ?
YES
NO
New CH$_4$/ΣC value
ELEMENTAL BALANCE
SO$_2$ ΔP CALC.
? ΔP = .3 P°
YES
NO
New Air value
HEAT BALANCE
? Close to Within .01%
YES
NO
New Fuel C/CaCO$_3$ Value
CO$_2$ ΔP
Meet Desired Value ?
YES
TYPE
NO
New Regn. Temp.

Program to calculate heat and elemental balances for production of low-sulfur fuel gas by the CO_2 Acceptor Process. BASIC programming language is used.

```
00040    DIM C(10,12),N(8,2),Q(10,3),W(6,2),F(6,3),S(6,3)
00041    DIM X(10,2),T(20),R(8,1),O(6,1),I(6,3)
00042    DIM J(6,3),V(6,4),Y(4,4),H(25),L(25)
00044    PRINT "NEED K'/K?? 1=YES,0=NO";
00046    INPUT Y9
00047    IF Y9=0 THEN 55
00048    PRINT "K'/K GASIF";
00050    INPUT X1
00055    PRINT "NEED INT TYPEOUT??,1=YES 0=NO";
00060    INPUT Y7
00061    PRINT "NEED FULL KIN?? 1=YES 0=NO";
00062    INPUT Y8
00065    PRINT TAB(5);"REGENERATOR"
00070    PRINT "%CO OUT";
00075    INPUT C7
00080    PRINT "%BURNOUT";
00085    INPUT C8
00100    PRINT "MOLS LIFT GAS";
00105    INPUT R5
00110    PRINT "LIFT GAS T";
00115    INPUT T(7)
00136    PRINT "NEED DP CALC?? 1=YES,0=NO";
00137    INPUT O1
00138    IF O1=0 THEN 141
00139    PRINT "DP CO2";
00140    INPUT P2
00141    PRINT
00145    PRINT "PRESS";
00150    INPUT P
00155    PRINT "GASIF T";
00160    INPUT T(1)
00165    PRINT "REGEN T";
00170    INPUT T(2)
00250    PRINT "BTU/LB";
00255    INPUT H4
00295    PRINT "INPUT SOLIDS"
00300    PRINT "H  ";
00305    INPUT H1
00310    PRINT "C  ";
00315    INPUT C0
00320    PRINT "N  ";
00325    INPUT N0
00330    PRINT "O  ";
00335    INPUT O0
00340    PRINT "S  ";
00345    INPUT S0
00350    PRINT "ASH";
00355    INPUT A9
00360    PRINT "%ACCEPT MAKEUP";
00365    INPUT M1
00366    IF P<5 THEN 380
00370    PRINT "ACCEPT ACTIV";
00375    INPUT X9
00380    PRINT "% IMP IN MGO.CAO";
```

```
00385   INPUT I1
00390   PRINT "% H IN GASIF CHAR";
00395   INPUT A3
00425   IF P<5 THEN 436
00430   PRINT "INPUT GAS T";
00435   INPUT T(3)
00436   PRINT TAB(5);"PREOXIDIZER"
00437   PRINT "% PREOX";
00438   INPUT W6
00439   PRINT "% O2 UTIL";
00440   INPUT W7
00443   PRINT "AIR T";
00444   INPUT T(9)
00447   T(8)=T(9)
00448   PRINT
00450   PRINT "INPUT H2O T";
00455   INPUT 1(6)
00460   IF P<5 THEN 470
00465   G(4)=.3
00466   G(5)=.7
00467   GOTO 525
00470   G(2)=.24
00475   G(3)=.23
00480   G(4)=.09
00485   G(5)=.34
00490   G(6)=.2
00525   PRINT
00530   PRINT "CH4/C";
00535   INPUT A1
00536   IF P>5 THEN 555
00538   B(1)=4E-5
00540   C=1E6
00542   I0=.3
00544   T(3)=T(1)
00546   X9=1E-5
00548   GOTO 630
00555   PRINT "CACO3/C";
00560   INPUT B(1)
00580   PRINT "LB FUEL C/CACO3";
00585   INPUT C
00620   PRINT "MOLS INLET GAS";
00625   INPUT I0
00630   PRINT "MOLS H2O TO GASIF";
00635   INPUT D1
00654   PRINT
00655   PRINT "HEAT LOSS GASIF";
00660   INPUT L(1)
00665   PRINT "HEAT LOSS REGEN";
00670   INPUT L2
00695   PRINT "% INPUT C REJECTED TO PRODUCT";
00700   INPUT Z3
00705   PRINT "NEED FULL TYPE?? 1=YES 0=NO";
00710   INPUT S9
00736   K5=EXP(71.5876-.03125*T(2)+5.79655E-6*T(2)↑2-67104/T(2))
00740   K4=EXP(17.5646-9.2926E-3*T(1)+1.7716E-6*T(1)↑2-1070/T(1))
00765   K1=EXP(-1.12894-9.3317E-4*T(1)+1.55385E-7*T(1)↑2+3239.62/T(1))
00770   I=1
00775   T=T(I)+460
00780   X(1,I)=1615*EXP(44.096-9.084E4/T)
00785   X(2,I)=18.52*EXP(29.179-6.011E4/T)
00790   X(3,I)=.05*EXP(31.548-6.499E4/T)
00795   X(4,I)=1.4*EXP(10.738-2.212E4/T)
00800   X(5,I)=41.8*EXP(39.587-8.155E4/T)
00805   X(6,I)=.491*EXP(4.6053-.9487E4/T)
00810   X(7,I)=.05*EXP(9.995-2.059E4/T)
00815   X(8,I)=.15*EXP(3.956E4/T-19.204)
```

```
00820   X(9,I)=1.4*EXP(9.7766-2.014E4/T)
00825   X(10,I)=.7*EXP(11.17-2.301E4/T)
00885   FOR I=1TO2
00890   T=(T(I)+460)/1.8
00895   P(I)=EXP((-41.5964+.0041409*T-.3279E-6*T↑2+43498/T)/-1.9872)
00896   NEXT I
00898   FOR I=0TO3
00900   C(1,I)=7.976+4.374E-3*T(I)-.4424E-6*T(I)↑2
00905   C(2,I)=6.868+.0045E-3*T(I)+.0045E-6*T(I)↑2
00910   C(3,I)=6.843+.4645E-3*T(I)-.0027E-6*T(I)↑2
00915   C(4,I)=8.891+2.283E-3*T(I)-.3579E-6*T(I)↑2
00920   C(5,I)=6.410+.5811E-3*T(I)-.007E-6*T(I)↑2
00921   NEXTI
00922   FOR I=1TO3
00925   C(6,I)=1026.53+.438232*T(I)+35.7628E-6*T(I)↑2+2.15368E-9*T(I)↑3
00930   C(7,I)=.207+.0317E-3*T(I)-.0034E-6*T(I)↑2
00935   C(8,I)=1.447+2.911E-3*T(I)-.643E-6*T(I)↑2
00940   C(9,I)=21.16+1.083E-3*T(I)
00945   C(10,I)=31.31+4.333E-3*T(I)
00950   NEXT I
00951   M(7)=9.274+.002798*T(2)-.898E-6*T(2)↑2+.121E-9*T(2)↑3
00952   M(8)=7.68+.001251*T(2)-.64E-6*T(2)↑2+.126E-9*T(2)↑3
00975   C1=1026.53+.438232*T(6)+35.7628E-6*T(6)↑2+2.15368E-9*T(6)↑3
00980   I=1
00982   R0=M1/(2/X9-2.34+1.17*M1)
00990   Z(1)=C0*Z3/1201
01035   G1=C0/12.01+(G(3)+G(4)+G(1))*I0
01055   G2=H1/2.016-.0705+(G(6)+G(2)+2*G(1))*I0-T3
01056   W8=W6/32
01057   IF W7=0 THEN 1060
01058   W9=W6/32*(100/W7-1)
01059   B4=11.76*W6/W7
01060   G3=O0/16+(G(3)+2*G(4)+G(6))*I0+2*(W8+W9)+T4
01095   A4=A1/((1-A1)*(1-B(I))-A1*B(I))
01096   F3=F3+1
01100   B1=1+B(I)*(1+C/12.01)/(1-B(I))
01105   G4=C*(1+A4)*B(I)*H3/(201.6*(1-B(I)))
01110   G9=(12.01*Z(1)+A9)*H3/201.6
01115   G5=G3+D1
01120   G6=G2+D1-G5
01125   G7=G4-2*B(I)*(1+A4)/(1-B(I))+2*A4
01130   G8=(G1-Z(1))/(B1*(1+A4))
01135   S1=G6-G7*G8+G8-G9
01140   S2=G5-G8-2*B(I)*(1+A4)*G8/(1-B(I))
01145   S3=S1+K1*(S2+G8)
01150   C9=K1*S2*G8
01155   IF ABS(1-K1)<.001 THEN 1185
01160   IF (S3↑2+4*C9*(1-K1))>6 THEN 1175
01165   PRINT I;"NEG"
01170   PAUSE
01175   N(4,I)=(-S3+SQR(S3↑2+4*C9*(1-K1)))/(2*(1-K1))
01180   GOTO 1190
01185   N(4,I)=C9/S3
01190   N(3,I)=G8-N(4,I)
01195   N(2,I)=G6-G7*G8+G8-G9+N(4,I)
01200   N(6,I)=G5-G8-N(4,I)-2*B(I)*(1+A4)*G8/(1-B(I))
01205   N(1,I)=A4*(N(4,I)+N(3,I))
01206   N(5,I)=G(5)*I0+B4
01207   T3=N(6,I)/K4
01208   T4=S0/32.06-T3
01209   T5=T4/(T3+T4)
01210   M(I)=(N(4,I)+N(3,I))*(1+A4)+N(2,I)+N(6,I)+N(5,I)+T3+.047
01215   IF P>5 THEN 1326
01220   I0=D1*M(I)/(S0*(N(3,I)+N(2,I))-N(6,I))
01225   FOR K=1TO6
01230   G(K)=N(K,I)/M(I)
```

```
01235   NEXT K
01326   O5=11.76*W6/W7+W8+W9
01327   O6=M3*O5/(1-M3)
01328   O7=H(8)*O6*18.02
01332   Q(3,1)=(18.02*(D1-O6)-6.383)*C1
01345   T=T(I)-60
01405   R(8,1)=0
01410   FOR K=1TO5
01415   R(K,1)=I0*G(K)*C(K,3)*(T(3)-60)
01420   R(8,1)=R(8,1)+R(K,1)
01425   NEXT K
01430   R(6,1)=G(6)*C(6,3)*18.02*I0
01435   R(8,1)=R(8,1)+R(6,1)
01436   B7=B4/.79*(T(8)-60)*(6.41+.5811E-3*T(8)-.007E-6*T(8)↑2)
01437   R(8,1)=R(8,1)+B7+O7
01540   Q(7,I)=A9*C(7,I)*T
01545   E(I)=0
01550   FOR K=1TO5
01555   S(K,I)=N(K,I)*C(K,I)*T
01560   E(I)=E(I)+S(K,I)
01565   NEXT K
01570   S(6,I)=N(6,I)*C(6,I)*18.02
01575   E(I)=E(I)+S(6,I)
01585   U(I)=(N(4,I)+N(3,I)+N(1,I))*B(I)/(1-B(I))
01590   I4=H3*(12.01*Z(1)+C*U(1)+A9)/(2.016*(100-H3))
01595   Q(8,I)=I4*C(2,I)*T
01615   Q(5,I)=(Z(1)+C*U(I)/12.01)*C(8,I)*T
01630   W(1,I)=N(3,I)*121760+N(1,I)*383020+N(2,I)*123180
01640   W(2,I)=(C*U(I)/12.01+Z(1))*(178540-4*T(I))+125190*I4
01670   W(1,I)=W(1,I)+7750+T3*242000
01675   W(5,I)=100*H4
01705   W(0,I)=W(1,I)+W(2,I)-W(5,I)
01720   B(0)=T4*196540
01722   B5=G(3)*I0*121760+G(2)*123180*I0+I0*G(1)*383020
01724   W(0,I)=W(0,I)-B5+B(0)
01725   W(6,I)=E(I)+W(0,I)+Q(5,I)+Q(7,I)+Q(8,I)+L(I)-R(8,I)-Q(3,I)
01745   W(6,I)=W(6,I)+769
01750   F(3,I)=C(9,2)*(T(2)-60)-C(9,I)*T
01755   I5=I1*96.4/(100-I1)
01760   F(5,I)=I5*(C(7,2)*(T(2)-60)-C(7,I)*T)
01765   F(1,I)=75215.8+1.16669*T(I)-.0025774*T(I)↑2
01770   F(0,I)=U(I)*((F(3,I)+F(5,I))/R0+F(1,I))-U(I)*T*C(4,I)
01775   Z1=(F(0,I)-W(6,I))/F(0,I)
01780   IF Y7=0 THEN 1818
01785   PRINT Z1;E(I);W(0,I);W(6,I);F(0,I);U(I);B(I)
01790   PRINT N(1,I);N(2,I);N(3,I);N(4,I);N(6,I);M(I);N(5,I)
01795   PRINT
01800   IF V<50 THEN 1818
01815   PAUSE
01818   IF Y>0 THEN 1820
01819   IF ABS(Z1)<.05 THEN 1955
01820   IF ABS(Z1)<.0002 THEN 1955
01825   IF V>50 THEN 1785
01835   IF ABS(Z1)<.5 THEN 1870
01840   B(I)=(1-Z1/5)*B(I)
01850   V=V+1
01855   GOTO 980
01870   B(I)=(1-Z1/2)*B(I)
01880   V=V+1
01885   GOTO 980
01955   V1=V1+V
01960   V=0
01961   IF Y9=1 THEN 8000
01965   GOTO 2305
01975   PRINT
01980   PRINT TAB(5);"GASIF"
```

```
01985   H2=P*N(6,1)/M(1)
01990   C3=P*N(4,1)/M(1)
01991   IF S9=0 THEN 2024
02000   K7=EXP(-5.41543-9.53865E-3*T(1)+1.39652E-6*T(1)↑2+16629/T(1))
02005   N7=N(1,1)*N(6,1)/N(3,1)/N(2,1)↑3*(M(1)/P)↑2
02010   PRINT "K'/K";DEC(N7/K7,10,2)
02011   PRINT "AI";DEC(A1,10,3);TAB(20);"CH4";DEC(N(1,1),26,4)
02015   PRINT DEC(B(1),12,4);TAB(20);"H2";DEC(N(2,1),26,4)
02016   Z7=N(1,1)+N(3,1)+N(4,1)+U(1)-I0*(G(3)+G(4))
02020   PRINT "CACO3/C";DEC(U(1)/Z7,12,4);TAB(20);"CO";DEC(N(3,1),26,4)
02021   PRINT "C GASIF";DEC(Z7,13,4);TAB(20);"CO2";DEC(N(4,1),26,4)
02022   PRINT "C TO REGEN";DEC((C*U(1)/12.01),13,4);TAB(20);"H2O";
02023   PRINT DEC(N(6,1),26,4)
02024   PRINT "%B";DEC(Z7*1201/C8,14,2);DEC((42-C*U(1))/.42,22,1)
02026   PRINT
02135   C4=P*N(2,1)/M(1)
02140   C5=P*N(3,1)/M(1)
02145   C6=P*N(1,1)/M(1)
02146   IF S9=0 THEN 7000
02150   R2=X(1,1)*(H2-C4*C5/X(2,1))/(1+X(3,1)*H2+.2*C4+X(4,1)*C5)↑2
02155   H(12)=X(5,1)*(H2*C4-C5*C6/X(6,1))
02156   R3=H(12)/(1+X(7,1)*H2+X(8,1)*C4+X(9,1)*C5+X(10,1)*C6)↑2
02160   PRINT "RC ";DEC(R2,14,1);TAB(20);"DELTA H";DEC(W(0,1),37,0)
02165   PRINT "RCH4 ";DEC(R3,14,1);TAB(20);"GAS SENS";DEC(E(1)+769,37,0)
02166   D0=W9*.79*P/(11.76*W6/W7)
02167   Y1=W9*P/(11.76*W6/W7/.79+D1)
02170   PRINT "MOLS GAS";DEC(M(1),14,3);TAB(20);"ACC DUTY";
02171   PRINT DEC(F(0,1),37,0);TAB(40);"BTU WET";
02172   PRINT DEC(W(1,1)/(M(1)*380),52,1)
02173   PRINT
02174   PRINT "MGO.CACO3";DEC(X,14,4)
02175   PRINT "%S REJ";DEC(100*T5,14,1);TAB(20);"PO2";DEC(D0,28,3);
02176   PRINT DEC(Y1,36,3)
02177   PRINT "CAS/MGO.CAO";DEC(T4*R0/X,14,4);
02178   PRINT TAB(20);"AIRN2";DEC(11.76*W6/W7,28,4)
02180   PAUSE
02305   T=T(2)-60
02315   T2=T(1)-60
02316   R4=0
02317   U(0)=0
02320   I6=(U(0)+U(1))/R0*M1/100
02330   Q1=(U(1)-I6*R0)*(C(10,2)*T-C(10,1)*T2)
02340   Q3=(U(1)/R0-I6)*I5*(C(7,2)*T-C(7,1)*T2)
02350   Q5=(U(1)*(1/R0-1)-I6*(1-R0))*(C(9,2)*T-C(9,1)*T2)
02355   Q6=(U(1)+I6)*(75215.8+1.16669*T(2)-.0025774*T(2)↑2)
02365   Q8=I6*(46000+C(10,2)*T)+I6*C(4,2)*T
02370   Q9=I6*I5*C(7,2)*T
02371   U2=(U(1))*C(4,2)*T
02372   U3=2*I6*C(4,2)*T
02375   Q7=Q9+Q8+Q6+Q5+Q3+Q1
02380   X=U(1)
02381   O4=T4*R0/X*I6
02382   T9=T4-O4
02383   S4=T9
02386   IF P>5 THEN 2390
02387   M3=.256/(P*14.7)
02388   GOTO 2395
02390   M3=.949/(P*14.7)
02395   E1=C8*U(1)*C/1201+U(1)+I6*(2-R0)
02400   E2=2*I6*(2-R0)+2*U(1)-3*T9-I4
02405   A2=C7/(100-C7)
02406   IF A2>0 THEN 2412
02408   L(15)=R5*(1-M3)/(1+1.88095/E1*(2*E1-E2))
02409   N(4,2)=E1+L(15)
02410   N(3,2)=0
02411   GOTO 2490
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02412   IF B>0 THEN 2420
02415   A=10
02422   E3=E1/(1+A)
02425   L(15)=R5*(1-M3)/(1/A+1+1.88095*(2+1/A-E2/A/E3))
02430   N(3,2)=E3+L(15)/A
02435   B2=((A*E3)+L(15))*(1+1.88095*(2+1/A-E2/(A*E3)))
02440   Z9=(B2-N(3,2)/A2)/B2
02455   IF ABS(Z9)<.00001 THEN 2485
02460   A=A*(1-Z9/2)
02465   B=B+1
02470   IF B<50 THEN 2420
02475   PRINT "B";Z9
02480   PAUSE
02485   N(4,2)=A*N(3,2)
02486   B=0
02490   L(12)=N(3,2)*L(15)/N(4,2)
02495   A3=(2*N(4,2)+N(3,2)-L(12)-2*L(15)-E2)/.42
02500   L(16)=(.79*A3*L(15))/(N(4,2)-L(15))
02505   N(6,2)=I4+R5*M3+A3*M3/(1-M3)
02510   N(5,2)=.79*A3+L(16)
02515   V3=N(3,2)+N(4,2)+N(5,2)+N(6,2)+T9
02520   L(13)=R5*M3
02541   IF P<5 THEN 2545
02542   GOTO 6000
02545   FOR I=7TO8
02550   H(I)=1039.25+.41916*T(I)+46.6E-6*T(I)↑2
02555   NEXT I
02556   I=1
02560   H(6)=1039.25+.41916*T(2)+46.6E-6*T(2)↑2
02565   H(9)=1021.08+.4854*T(9)-35.83E-6*T(9)↑2+34.46E-9*T(9)↑3
02570   H(5)=6.925+1.967E-4*T(9)+1.059E-7*T(9)↑2
02575   V5=C*U(1)/12.01*(178540-4*T(1))+125190*I4
02580   V6=(1-C8/100)*U(1)*C/12.01*(178540-4*T(2))
02585   V4=N(3,2)*121760
02590   U1=L(12)*121760
02595   IF P<5 THEN 2620
02596   L(15)=(T8*(1+A)+S5-C*U(1)*C8/1201-X-I6*(2-R0))/(1+1/A)
02600   L(12)=L(15)/A
02602   L(16)=L(15)*N(5,2)/A/T8
02604   L(0)=T8+2*A*T8+2*S4+T6+S5-2*I6*(2-R0)
02605   A3=(L(0)+T9-2*U(1)-L(12)-2*L(15)-L(13))/(.42+M3/(1-M3))
02606   N(5,2)=.79*A3+L(16)
02607   N(6,2)=M3*A3/(1-M3)+M3*R5-H0-S7+I4
02608   R7=(N(3,2)+2*(N(4,2)+S4)+N(6,2)+S5+T9)
02609   B3=A3*(.42+M3/(1-M3))+2*I6*(2-R0)+2*U(1)+L(12)+2*L(15)+L(13)-R7
02610   V3=H0+T8*(1+A)+N(6,2)+.79*A3+L(16)+S4+S5+S6+S7
02611   B4=S4+2*S6
02612   V4=N(3,2)*121760+N(2,2)*123180+S7*23800+S5*242000
02613   S4=S4+B3/2
02614   U1=L(12)*121760
02615   S6=(B4-S4)/2
02617   U9=S4*196540+S6*82120
02618   C7=T8/(V3-T6-.7*T9)*100
02620   M4=M3*A3/(1-M3)*18.02*H(9)
02622   M6=M3*R5*18.02*H(7)
02624   V7=Q(5,1)+Q(7,1)+Q(8,1)
02626   M7=M4+M6
02628   U6=A3*H(5)*(T(9)-60)
02635   C(3,7)=6.843+.4645E-3*T(7)-.0027E-6*T(7)↑2
02640   C(4,7)=8.891+2.283E-3*T(7)-.3579E-6*T(7)↑2
02645   C(5,7)=6.41+.5811E-3*T(7)-.007E-6*T(7)↑2
02646   IF P<5 THEN 2650
02647   U8=(T(7)-60)*((L(12)+L(16))*C(5,7)+L(15)*C(4,7))
02648   GOTO 2652
02650   H(18)=R5*(1-M3)*(T(7)-60)/(V3*(1-M3))
02651   U8=H(18)*(N(3,2)*C(3,7)+N(4,2)*C(4,7)+N(5,2)*C(5,7))
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02652   IF P>5 THEN 2680
02654   U9=T9*196500
02680   V8=A9*C(7,2)*T
02685   U7=(1-C8/100)*U(1)*C/12.01*C(8,2)*T
02690   M9=0
02695   FOR K=2TO5
02700   M(K)=N(K,2)*C(K,2)*T
02705   M9=M9+M(K)
02710   NEXT K
02715   M(6)=N(6,2)*H(6)*18.02
02716   IF P>5 THEN 2720
02717   M9=M9+M(6)
02718   GOTO 2725
02720   M9=M9+M(6)+T*(M(7)*S4+M(8)*S6)
02725   U0=M9+U7+Q7+L2-U8-U6-M7-V7+V8-U2-U3
02730   H6=V5+U1-V6-V4+U9
02735   Z4=(U0-H6)/U0
02736   F4=F4+1
02737   G=0
02740   IF Y7=0 THEN 2755
02745   PRINT Z4;M9;U0;H6;Q7;C
02750   PRINT N(3,2);N(4,2);N(5,2);N(6,2);V3
02751   PAUSE
02755   IF Y< 50 THEN 2765
02760   PAUSE
02765   IF ABS(Z4)<.0001 THEN 2810
02770   IF Y>50 THEN 2745
02775   IF ABS(Z4)<.2 THEN 2795
02780   C=(1+Z4/2)*C
02785   Y=Y+1
02786   GOTO 4000
02795   C=(1+Z4)*C
02800   Y=Y+1
02801   GOTO 4000
02810   Y=1
02812   IF Y7=1 THEN 2985
02822   IF O1=0 THEN 2826
02824   GOTO 8140
02826   PRINT "V1";V1;"F1";F1;"H";H;"S";S
02827   PRINT "ITR";F3+F4;"GASIF";F3;"REGEN";F4
02828   V1=Y=F3=F4=H=S=F1=0
02840   IF S9=0 THEN 1985
02841   PRINT TAB(5);"REGEN"
02845   W5=N(4,2)*P/V3
02850   PRINT "C";DEC(C,5,3);TAB(20);"DUTY";DEC(H6,32,0)
02851   O2=T8/(V3-T6)*100
02855   PRINT "DP CO2";DEC((P(2)-W5),12,3);TAB(20);"%CO";DEC(O2,26,2)
02880   PRINT "MGO.CAO";DEC(X/R0,14,3)
02890   PRINT "MOLS GAS";DEC(V3,14,3);TAB(20);"DP SO2";
02891   PRINT DEC(K5-S4*P/V3,30,4)
02912   Y6=N(1,1)*383020
02914   N2=N(3,1)*121760
02916   N3=N(2,1)*123180
02918   H(21)=Y6+N2+N3+7750+T3*242000
02934   PRINT "T";DEC(T(2),17,0)
02935   PRINT "MOLS AIR";DEC(A3,14,4);TAB(20);"HT/DUTY";DEC(U0/Q7,30,3)
02936   PRINT
02950   PRINT "EFF";DEC((H(21)/(100*H4)),14,4)
02951   IF D1<1 THEN 2970
02955   PRINT "STEAM CONV";
02960   N4=(D1-N(6,1))/D1
02965   PRINT DEC(N4,14,4);TAB(20);"R0";DEC(R0,30,4)
02970   N8=X*M1*(184.4+1.456*I5)/(100*R0)
02975   PRINT "LB STONE";DEC(N8,14,2);
02976   B4=162500*N8/W(1,1)
02977   PRINT TAB(20);"C/MM";DEC(B4,30,2)
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02980   GOTO 1975
02985   PAUSE
04000   IF P<5 THEN 980
04002   X5=2*N(4,2)+N(3,2)+N(6,2)+2*S4-2*L(15)-L(12)-L(13)
04004   X6=N(6,2)+H0+S7-L(13)
04006   X7=N(4,2)+N(3,2)+S5-L(15)-L(12)
04008   N(2,2)=.67*H0
04010   N(6,2)=X6-N(2,2)
04012   N(4,2)=X5-N(6,2)-X7
04014   N(3,2)=X7-N(4,2)
04016   I0=N(2,2)+N(3,2)+N(4,2)+N(5,2)+N(6,2)-L(16)
04017   N(5,2)=N(5,2)-L(16)
04018   FOR K=2TO6
04020   G(K)=N(K,2)/I0
04022   NEXT K
04024   GOTO 980
04070   N(2,2)=H0
04075   N(3,2)=T8
04080   N(4,2)=A*T8
04085   N(6,2)=T6
04105   GOTO 2545
06000   I=0
06005   T7=N(3,2)+N(4,2)
06010   O=N(3,2)+2*N(4,2)+N(6,2)+2*T9
06015   D=N(6,2)
06020   IF S>0 THEN 6030
06025   W=T7+D+N(5,2)+.6*T9
06030   K=0
06035   T1=(T(2)+460)/1.8
06040   K8=EXP(.278777-2.73705E-4*T1+4.37722E-8*T1↑2-11554.5/T1)
06045   K2=EXP(-6.24562+1.71757E-3*T1-3.52622E-7*T1↑2+5198/T1)
06050   K3=EXP(17.5646-9.2926E-3*T(2)+1.7716E-6*T(2)↑2-1070/T(2))
06055   K9=EXP(-1.4871-2.16103E-3*T1+4.39584E-7*T1↑2+14786/T1)
06056   K5=EXP(71.5076-.03125*T(2)+5.79655E-6*T(2)↑2-67104/T(2))
06060   IF S>0 THEN 6070
06065   A=10
06070   B1=1+2*K8
06075   G=G+1
06080   B2=2+2*K8
06085   B3=1+2/K3
06090   T6=A*D/(K2+A*B3)
06095   B4=(1+A*B2)*T7/(1+A*B1)
06100   S4=(O-T6-B4)/2
06105   IF S4>0 THEN 6120
06110   A=A/1.1
06115   GOTO 6090
06120   S6=(K9*S4)↑2*((D-T6*B3)/T6)↑4*P/W
06125   S1=2*S6+2*K8*A*T7/(1+A*B1)+2*T6/K3+S4
06130   Z=(T9-S1)/T9
06135   IF ABS(Z)<.001 THEN 6265
06136   IF ABS(Z)<.01 THEN 6261
06140   IF ABS(Z)<.2 THEN 6220
06145   IF Z>0 THEN 6160
06150   IF (Z+2)>0 THEN 6220
06155   IF Z<0 THEN 6180
06160   A=.99*A
06165   K=K+1
06170   IF K<100 THEN 6090
06175   GOTO 6195
06180   A=1.014*A
06185   K=K+1
06190   IF K<100 THEN 6090
06195   PRINT "K";K;"I";I;Z;A;S4;S6
06200   PAUSE
06205   I=0
06210   K=0
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06215   GOTO6090
06220   I=I+1
06225   W=T7+D+N(5,2)+S4+S6
06230   IF Z<0 THEN 6250
06235   A=.995*A
06240   IF I<100 THEN 6090
06245   GOTO 6195
06250   A=1.006*A
06255   IF I<100 THEN 6090
06260   GOTO 6195
06261   A=(1-Z/20)*A
06262   GOTO 6090
06265   T8=T7/(1+A*B1)
06270   H0=D-T6*B3
06275   S5=2*K8*A*T8
06280   S7=2*T6/K3
06285   E=K5-S4/W*P
06290   E6=.3*K5
06295   Z6=(E6-E)/E6
06300   IF ABS(Z6)<.05 THEN 4070
06305   IF ABS(Z6)<.3 THEN 6340
06310   Y3=O*(1-Z6/100)
06312   GOTO 6342
06325   PRINT "O";O;"Z6";Z6
06330   G=0
06335   PAUSE
06340   Y3=O*(1-Z6/50)
06342   N(5,2)=1.881*(Y3-O)+N(5,2)
06344   O=Y3
06346   W=T7+D+N(5,2)+S4+S6
06348   S=S+1
06350   IF G<20 THEN 6075
06355   GOTO 6325
07000   J0=N(1,1)+N(3,1)+N(4,1)+U(1)-I0*(G(3)+G(4))
07005   J1=(G2-G(6)*I0-I4)/4
07010   J2=(G3-I0*(G(3)+2*G(4)+G(6))-2*(W8+W9))/4
07015   J3=A1/4
07020   J4=J0/4
07025   FOR K=3TO1 STEP-1
07030   I(1,K)=J3*J4*K↑2
07035   L3=K*J4*(1-B(1))-I(1,K)
07036   L3=L3+I0*(G(3)+G(4))
07040   L4=D1+K*J1-2*I(1,K)
07041   L4=L4+I0*G(6)
07045   L5=D1+K*J2-2*K*J4*B(1)
07046   H(23)=I0*(G(3)+2*G(4)+G(6))+2*(W8+W9)
07047   L5=L5+H(23)
07050   L6=L5-L4-L3
07055   L7=L3-L6
07060   L8=L7*L4
07065   L9=L7+L4
07070   J6=K1*L8
07075   J5=L6+K1*L9
07080   J7=K1-1
07085   IF ABS(J7)<.001 THEN 7100
07090   I(2,K)=(J5-SQR(J5↑2-4*J6*J7))/(2*J7)
07095   GOTO 7105
07100   I(2,K)=J6/J5
07105   I(4,K)=I(2,K)+L6
07110   I(3,K)=L3-I(4,K)
07115   I(6,K)=L4-I(2,K)
07120   H(K)=I(1,K)+I(2,K)+I(3,K)+I(4,K)+I(6,K)+N(5,1)
07125   FOR L=1TO 6
07130   J(L,K)=I(L,K)/H(K)*P
07135   NEXT L
07140   NEXT K
```

```
07141   PRINT
07142   IF S9=1 THEN 7146
07143   Y8=0
07146   Z8=D1+I0+11.76*W6/W7
07147   J(6,0)=(D1+I0*G(6))/Z8*P
07148   FOR L=1TO4
07149   J(L,0)=I0*G(L)*P/Z8
07150   NEXT L
07152   PRINT "H20";DEC(H2,8,3);
07154   IF Y8=0 THEN 7175
07155   FOR K=3TO0 STEP-1
07165   PRINT DEC(J(6,K),(40-8*K),3);
07170   NEXT K
07175   PRINT
07180   PRINT "H2";DEC(C4,8,3);
07181   IF Y8=0 THEN 7200
07185   FOR K=3TO0 STEP-1
07190   PRINT DEC(J(2,K),(40-8*K),3);
07195   NEXT K
07200   PRINT
07205   PRINT "CO";DEC(C5,8,3);
07206   IF Y8=0 THEN 7225
07210   FOR K=3TO0 STEP-1
07215   PRINT DEC(J(3,K),(40-8*K),3);
07220   NEXT K
07225   PRINT
07230   PRINT "CO2";DEC(C3,8,3);
07231   IF Y8=0 THEN 7250
07235   FOR K=3TO0 STEP-1
07240   PRINT DEC(J(4,K),(40-8*K),3);
07245   NEXT K
07250   PRINT
07251   IF P<5 THEN 7285
07255   PRINT "DP CO2";DEC((C3-P(1)),10,3);
07260   FOR K=3TO0 STEP-1
07265   PRINT DEC((J(4,K)-P(1)),(40-8*K),3);
07270   NEXT K
07275   PRINT
07285   V(6,4)=.9*H2+.1*J(6,3)
07290   V(2,4)=.9*C4+.1*J(2,3)
07295   V(3,4)=.9*C5+.1*J(3,3)
07296   V(4,4)=.9*C3+.1*J(4,3)
07300   V(1,4)=.9*C6+.1*J(1,3)
07305   FOR K=3TO2 STEP-1
07310   FOR L=1TO6
07315   V(L,K)=.9*J(L,K)+.1*J(L,(K-1))
07320   NEXT L
07325   NEXT K
07330   V(6,1)=.9*J(6,1)+.1*(D1+I0*G(6))/Z8*P
07335   FOR L=1TO 3
07340   V(L,1)=.9*J(L,1)+.1*(I0*G(L))*P/Z8
07341   NEXT L
07345   V(4,1)=.9*J(4,1)+.1*(I0*G(4))*P/Z8
07350   FOR L=4TO1 STEP-1
07355   H(19)=X(1,1)*(V(6,L)-V(3,L)*V(2,L)/X(2,1))
07356   Y(1,L)=H(19)/(1+X(3,1)*V(6,L)+.2*V(2,L)+X(4,1)*V(3,L))↑2
07360   H(20)=X(5,1)*(V(6,L)*V(2,L)-V(3,L)*V(1,L)/X(6,1))
07361   H(22)=1+X(7,1)*V(6,L)+X(8,1)*V(2,L)
07362   Y(2,L)=H(20)/(H(22)+X(9,1)*V(3,L)+X(10,1)*V(1,L))↑2
07365   NEXT L
07425   FOR L=4TO1 STEP-1
07430   Y(3,L)=Y(1,L)+2*Y(2,L)
07440   NEXT L
07475   J8=1/Y(3,4)+1/Y(3,3)+1/Y(3,2)+1/Y(3,1)
07480   PRINT "R BAR";DEC(4/J8,12,1)
07483   Y8=1
```

```
07485    GOTO 2985
08000    K7=EXP(-5.41543-9.53865E-3*T(1)+1.39652E-6*T(1)↑2+16629/T(1))
08005    N7=N(1,1)*N(6,1)/N(3,1)/N(2,1)↑3*(M(1)/P)↑2
08010    X2=N7/K7
08015    P1=(X1-X2)/X1
08020    IF ABS(P1)<.01 THEN 8050
08025    A1=A1*(1+P1/2)
08026    IF A1>0 THEN 8030
08027    A1=.05
08030    F=F+1
08035    IF F<20 THEN 980
08040    PRINT "F";"A1";A1
08045    PAUSE
08050    F1=F1+F
08055    F=0
08130    GOTO 2305
08140    W5=N(4,2)/V3*P
08145    P4=P(2)-W5
08150    P5=(P2-P4)/P2
08155    IF ABS(P5)<.01 THEN 2826
08160    IF P5<0 THEN 8190
08165    T(2)=T(2)+20*P5
08170    H=H+1
08175    IF H<20 THEN 885
08180    PRINT "H";DEC(T(2),8,0)
08185    PAUSE
08190    T(2)=T(2)+28*P5
08195    H=H+1
08200    IF H<20 THEN 885
08205    GOTO 8180
09000    PRINT TAB(30);"REGEN"
09004    PRINT TAB(37);"HT. CONTENT"
09005    PRINT TAB(22); "MOLS";TAB(37);"OR DELTA H"
09010    PRINT TAB(5);"[IN]"
09015    PRINT "CHAR"
09020    PRINT "  C";DEC(C*U(1)/12.01,22,4);DEC(Q(5,1),39,0)
09025    PRINT"  H";DEC(I4,22,4);DEC(Q(8,1),39,0)
09030    PRINT "  ASH";DEC(Q(7,1),39,0)
09035    PRINT " TOTAL";DEC(V7,46,0)
09036    PRINT
09040    PRINT "DRY AIR";DEC(T(9),13,0);"F"DEC(A3,22,4);DEC(U6,39,0)
09045    PRINT "  H2O";DEC(M3*A3/(1-M3),22,4);DEC(M4,39,0)
09046    PRINT " TOTAL";DEC((U6+M4),46,0)
09050    PRINT "LIFT GAS"
09060    PRINT "  CO2";DEC(L(15),22,4)
09065    PRINT "  CO";DEC(L(12),22,4)
09070    PRINT "  N2";DEC(L(16),22,4);DEC(U8,39,0)
09075    PRINT "  H2O";DEC(L(13),22,4);DEC(M6,39,0)
09076    N5=L(12)+L(13)+L(15)+L(16)
09080    PRINT " TOTAL";DEC(N5,22,4);DEC(R5,28,4);DEC(U8+M6,46,0)
09096    PRINT "TOTAL IN";DEC((U8+U6+M7+V7),52,0)
09097    PRINT
09098    PRINT TAB(5);"[OUT]"
09100    PRINT "ACCEPTOR DUTY"
09105    PRINT "  MGO.CAO";DEC(X/R0,22,4)
09111    PRINT "  INERT(LB)";DEC(I5*X/R0,24,2)
09115    F7=U(1)*(C(10,2)*T-C(10,1)*T2)
09120    F8=(U(1)*(1/R0-1))*(C(9,2)*T-C(9,1)*T2)
09125    PRINT "  SENSIBLE MG.CA";DEC((F7+F8+Q8+Q4),39,0)
09130    F9=I5/R0*(U(1)*(C(7,2)*T-C(7,1)*T2)+U(0)*(C(7,2)*T-C(7,0)*T1))
09135    PRINT "  SENSIBLE INERT";DEC(F9,39,0)
09140    PRINT "  HT OF REACT";DEC(X*Q6/(X+I6),39,0)
09145    PRINT
09150    PRINT "  MAKEUP";DEC(I6,22,4)
09155    PRINT "  INERT(LB)";DEC(I5*I6,24,2)
09160    I8=I6*Q6/(X+I6)
```

```
09165   PRINT "  SENSIBLE+HT REACT";DEC((I8+Q8+Q9),39,0)
09170   PRINT
09175   PRINT " REJECT"
09180   PRINT "  MGO.CAO";DEC(I6*(1-R0)-O4,22,4)
09185   PRINT "  MGO.CACO3";DEC(I6*R0,22,4)
09186   PRINT "  MGO.CAS";DEC(O4,22,4)
09190   PRINT "  SENSIBLE";DEC(-(F7+F8+F9)*M1/100,39,0)
09200   PRINT " TOTAL ACCEPTOR DUTY";DEC(Q7,46,0)
09210   PRINT "CHAR"
09215   PRINT "  C";DEC(((1-C8/100)*C*U(1)/12.01),22,4);DEC(U7,39,0)
09220   PRINT "  ASH";DEC(V8,39,0)
09225   PRINT " TOTAL";DEC((U7+V8),46,0)
09230   PRINT "GAS"
09235   PRINT "  CO";DEC(N(3,2),22,4)
09240   PRINT "  CO2";DEC(N(4,2),22,4)
09245   PRINT "  N2";DEC(N(5,2),22,4)
09250   PRINT "  H2O";DEC(N(6,2),22,4)
09251   PRINT "  H2";DEC(H0,22,4)
09252   PRINT "  SO2";DEC(S4,22,4)
09253   IF P<5 THEN 9257
09254   PRINT "  S2";DEC(S6,22,4)
09255   PRINT "  H2S";DEC(S7,22,4)
09256   PRINT "  COS";DEC(S5,22,4)
09257   PRINT " TOTAL";DEC(V3,28,4);DEC(M9,46,0)
09261   PRINT "HT CONTENT OF CO2"
09262   PRINT "(MGO.CACO3+MAKEUP)";DEC(-(U2+U3),46,0)
09275   PRINT "LOSS";DEC(L2,46,0)
09280   PRINT "TOTAL OUT";DEC((M9+U7+V8+Q7+L2-U2-U3),52,0)
09285   PRINT "NET DUTY";DEC(U0,52,0)
09290   PRINT "HT. OF COMB."
09295   PRINT " CHAR IN";DEC(V5,46,0)
09300   PRINT " CO IN";DEC(U1,46,0)
09305   PRINT " CHAR OUT";DEC(-V6,46,0)
09310   PRINT " GAS OUT";DEC(-V4,46,0)
09311   PRINT "S REACTIONS";DEC(U9,46,0)
09315   PRINT "NET HEAT SUPPLIED";DEC(H6,52,0)
09320   PRINT TAB(30);"GASIF"
09335   PRINT TAB(5);"[IN]"
09345   PRINT "COAL"
09350   PRINT "  C";DEC(C0/12.01,22,4)
09355   PRINT "  H";DEC(H1/2.016,22,4)
09360   PRINT "  O";DEC(O0/16,22,4)
09365   PRINT "  ASH(LB)";DEC(A9,24,2)
09425   PRINT "STEAM";DEC(D1-O6-6.383/18.02,22,4);DEC(Q(3,1),46,0)
09426   PRINT " MOIST";DEC(.3542,22,4)
09427   IF P>5 THEN 9435
09428   PRINT "REDUCTANT";DEC(I0,22,4)
09430   GOTO 9465
09435   PRINT "CLAUS GAS"
09440   PRINT "  C0";DEC(G(3)*I0,22,4)
09445   PRINT "  CO2";DEC(G(4)*I0,22,4)
09450   PRINT "  N2";DEC(G(5)*I0,22,4)
09455   PRINT "  H2O";DEC(G(6)*I0,22,4)
09456   PRINT "  H2";DEC(G(2)*I0,22,4)
09460   PRINT " TOTAL SENS";DEC((R(8,1)-B7-O7),46,0)
09465   PRINT "AIR"
09470   PRINT "  N2";DEC(11.76*W6/W7,22,4)
09475   PRINT "  O2";DEC(W8+W9,22,4)
09476   PRINT "  H2O";DEC(O6,22,4);DEC(O7,39,0)
09480   PRINT " TOTAL";DEC(B7+O7,46,0)
09510   PRINT
09515   PRINT "TOTAL IN";DEC(R(8,1)+Q(3,1),52,0)
09520   PRINT
09525   PRINT TAB(5);"[OUT]"
09530   PRINT "CHAR"
09535   PRINT "  C";DEC((C*U(1)/12.01+Z(1)),22,4);DEC(Q(5,1),39,0)
```

```
09540    PRINT "  H";DEC(I4,22,4);DEC(Q(8,1),39,0)
09545    PRINT "  ASH(LB)";DEC(A9,24,2);DEC(Q(7,1),39,0)
09550    PRINT " TOTAL";DEC(Q(5,1)+Q(8,1)+Q(7,1),46,0)
09555    PRINT "GAS"
09560    PRINT "  CH4";DEC(N(1,1),22,4)
09565    PRINT "  H2";DEC(N(2,1),22,4)
09570    PRINT "  CO";DEC(N(3,1),22,4)
09575    PRINT "  CO2";DEC(N(4,1),22,4)
09580    PRINT "  H2O";DEC(N(6,1),22,4)
09590    PRINT "  N2";DEC(N(5,1),22,4)
09600    PRINT "  NH3";DEC(.047,22,4)
09601    PRINT "  H2S";DEC(T3,22,4)
09605    PRINT " TOTAL";DEC(M(1),28,4);DEC(E(1),46,0)
09625    PRINT "HT. OF REACT"
09630    PRINT "  CHAR OUT";DEC(W(2,1),39,0)
09640    PRINT "  GAS OUT";DEC(W(1,1),39,0)
09645    PRINT "  COAL IN";DEC(-W(5,1),39,0)
09646    IF P>5 THEN 9655
09648    PRINT "  REDUCTANT";DEC(-B5,39,0)
09650    GOTO 9660
09655    PRINT "  CLAUS GAS";DEC(-B5,39,0)
09660    PRINT " S REACTIONS";DEC(B(0),39,0)
09665    PRINT " NET HT OF REACT";DEC(W(0,1),46,0)
09670    PRINT "LOSS";DEC(L(1),46,0)
09680    L(20)=E(1)+W(0,1)+R(7,1)+Q(5,1)+Q(7,1)+Q(8,1)+L(1)
09695    PRINT "TOTAL OUT";DEC(L(20),52,0)
09705    PRINT "NET DUTY";DEC(W(6,1),52,0)
09710    PRINT "ACCEPTOR DUTY"
09715    PRINT "  SENSIBLE MGO.CAO";DEC(U(1)/R0*F(3,1),39,0)
09720    PRINT "  SENSIBLE INERT";DEC(U(1)/R0*F(5,1),39,0)
09725    PRINT "  HT OF REACT";DEC(U(1)*F(1,1),39,0)
09730    PRINT "  CO2 SENS HT";DEC(-U(1)*(T(1)-60)*C(4,1),39,0)
09735    PRINT " NET ACCEPTOR DUTY";DEC(F(0,1),52,0)
09745    PRINT "  MGO.CACO3 OUT";DEC(U(1),22,4)
09746    PRINT "  MGO.CAS OUT";DEC(T4,22,4)
09755    GOTO 2985
09760    END
*
```

APPENDIX E

Literature Review on Preoxidation of Caking Coals

A. Introduction and Summary of Prior Consol Work

Because of its specific purpose, the Bibliography for this Appendix has been kept separate (as Section C) from the main Bibliography for this Report.

The gasification process under development for OAP calls for the use of Eastern bituminous coals which are highly caking. When fed into the gasifier without pretreatment, these coals agglomerate, making operation of the fluidized bed very difficult.

A fluidized bed coal gasification pilot plant was operated in 1949 at the Research Division of Consolidation Coal Company in a joint program with Standard Oil Development Company.(1) It was demonstrated in the course of this work that Pittsburgh seam coal could be successfully fed to an air-blown fluidized bed operated at 1750°F without pretreatment.

These results, however, do not show that operation without pretreatment is feasible for the presently envisaged adaptation of the CO_2 Acceptor Gasification Process. The demonstrated coal feed rate, for example, was 20 $lb/hr\text{-}ft^2$ of gasifier cross section which is much too low for the present application. The above gasification work was also carried out at atmospheric pressure. Subsequent work indicates that the need for pretreatment increases with the operating pressure and particularly as the hydrogen partial pressure increases.

Work carried out by Consol for the Office of Coal Research in a continuous bench-scale unit used for development of the CO_2 Acceptor Process(2) showed, for example, that mild preoxidation (4%) was required to prevent coking in fluid bed hydro devolatilization of subbituminous coals at 1500°F and 20 atm pressure (H_2 partial pressure ca. 5.5 atm). No such pretreatment is required in the corresponding operation at atmospheric pressure.

The nature of the atmosphere at the point of injection of the coal also affects the coking tendency and need for pretreatment. The same subbituminous coal could be fed successfully without coking to the identical gasifier vessel operated at 1775°F and 20 atm pressure as a "CO maker."(3) The principal difference between this operation and the previous hydrodevolatilization is the relatively low partial pressure of hydrogen existing at the coal inlet. Thus, it is concluded that a highly reducing atmosphere increases the need for pretreatment.

Caking bituminous coals in contrast to the subbituminous coals were found, however, to require a high degree of preoxidation when fed under the same conditions to the same gasifier operated as a "CO maker." In fact even when 13% preoxidized coal was used* some caking was observed in the gasifier. The high degree of pretreatment required in this case is associated both with the high operating pressure and the reducing environment existing in the gasifier at the point of coal injection.

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* % preoxidation = 100(lb O_2 reacted)/lb dry coal.

Experimental evidence indicates, however, that the degree of pretreatment required to sustain operability is reduced when the coal is delivered into the gasifier in an oxidizing atmosphere. This is the method of operation proposed in the present adaptation of the CO_2 Acceptor Process. This is thought to be the reason why the air blown fluidized bed gas producer operation mentioned previously(1) could be carried out without pretreatment, even at the low rates demonstrated.

Similarly, highly caking Pittsburgh seam coals were used successfully as fuels with only mild preoxidation (5.3 wt %) for fluidized bed reduction of sulfated dolomite at temperatures up to 1875°F at a pressure of 8 psig.(4) The success in this instance is likely due to the highly oxidizing atmosphere existing at the point of injection of the coal into the bed. In this instance, straight air was used as transport gas to carry the coal into the bed.

The preoxidation process itself was studied extensively by the Research Division of Consolidation Coal Company in the 1950's. The objective of the work was to use preoxidation as a pretreatment method to render the coal operable when fed to a fluidized bed carbonization process operated in the temperature range of 900-950°F.

A comprehensive survey of the preoxidation literature prior to 1952 was made(5) at that time. A survey of the preoxidation literature since 1952 was made under the OAP contract and this is given here. It concentrates on work carried out in the temperature range of 600-1000°F. The reason for this will be evident in what follows. Some of the highlights of the previous literature(5) search are also given.

The earlier Consol laboratory program on preoxidation was conducted in three continuous bench-scale units with preoxidation vessels having internal diameters of 1-1/2",(6) 4",(7) and 6",(8) respectively.

The laboratory study was made in the temperature range of 650-825°F. The study included a determination of reaction kinetics, heat of reaction, effect of process variables such as particle size and determination of the effect of preoxidation conditions on properties of the product. The latter included determination of caking properties, bulk densities and other properties of the preoxidized coals. The general findings are discussed later in this survey in conjunction with a comparison to published literature data.

The main thrust of the program was, however, directed towards maximizing tar yield in the overall process. The approach used was to minimize the amount of preoxidation required for operability in subsequent fluid bed carbonization. This objective is achieved by combining preoxidation with thermal pretreatment, i.e., by operating at temperatures above 650°F wherein pyrolysis of the coal occurs simultaneously with preoxidation. The maximum temperature that can be employed is dictated by the operability limits imposed by the coal itself, since the preoxidation step usually must be operated below the point at which the "untreated" coal attains its maximum fluidity. In most cases, depending on the caking properties of the coal employed, the preoxidation temperature thus must be below 825°F. With this method an "optimum" preoxidation temperature is defined which corresponds to the minimum oxygen input to the preoxidizer required to maintain operability in both the preoxidation and carbonization steps. The optimum temperature is a function of the particular coal employed but is generally within the range of 700-800°F.

The above principle of an optimum preoxidation temperature in the plastic range of the coal was successfully demonstrated in a 30 TPD pilot plant. It is outside the scope of this survey to discuss the operation of the pilot plant. A good description of the process as finally developed is given, however, in a U. S. Patent granted to Consolidation Coal Company.(9)

The above background data on preoxidation forms a useful guide to our impending experimental program for OAP. It minimizes but does not eliminate further experimental work required to satisfy our present objective. It is likely that in this case also it will be desirable to operate the preoxidizer within the plastic range of the coal, i.e., 700-800°F. It is, however, no longer necessary to minimize the amount of preoxidation as in the case of the low-temperature carbonization development. It is merely sufficient not to exceed the oxygen input as required by heat balance to sustain temperature in an adiabatic preoxidizer. The operability limits in the preoxidizer itself are affected by operation under pressure and will thus have to be determined separately. Likewise, the extent of pretreatment required for the gasification operation and how it is affected by the method of introducing the coal into the gasifier still remains to be determined experimentally.

Finally, an important new consideration here is the desirability of producing a relatively high density char for the gasifier operation. This is a requirement which likely calls for maximization of the amount of preoxidation. Practical considerations, as discussed above, limit the amount of preoxidation to that which is consistent with the heat balance in the preoxidizer.

B. Discussion of Literature

1. Oxidation Kinetics

a. Mechanism of Oxidation

The oxidation of coal is extremely complicated. Depending upon the severity of reaction conditions, every organic oxidation product from peroxides to water and carbon dioxide may be observed. An exact model would involve taking into account a number of simultaneous and consecutive reactions. The reaction progress is generally evaluated using the amount of oxygen reacted per pound of coal fed or the fraction of unreacted material remaining in the coal.

b. Dependence of Rate Upon Degree of Coal Conversion

Struck(6) investigated kinetics of the oxidation of Disco char, Montour No. 10 coal, and Arkwright coal from about 550 to 765°F. Conversion was expressed as oxygen consumed, wt % of MAF coal. At 650°F, a log-log plot of rate vs. oxygen consumed was linear up to 20%. The slope was about -1, thus indicating the rate to be inversely proportional to the extent of reaction.

Weisman(10) studied the rate of dehydrogenation of bituminous coal during oxidation in the range of 300-710°F. He found the rate to be first order in unreacted hydrogen remaining in the coal up to about 20% loss of the hydrogen in the coal. The rate expression then became more complex as secondary reactions attained importance.

One may expect the reaction rate to be proportional to the material remaining available for reaction. The problem is what basis to use for the reactive material. Weisman uses the total hydrogen but divides it into more reactive and less reactive fractions. Struck used the amount of oxygen consumed to follow the reaction progress. Still another method is to measure the SCF oxygen reacted per pound of coal. One percent oxidation equals about 0.12 SCF/lb coal.

c. Effect of Oxygen Concentration

In his literature survey,(5) Struck found the rate of oxidation below 400°F to be correlated by $(P_{O_2})^h$ where h was anywhere from 0.5 to 0.7. Struck assumed, however, a first order relationship with respect to oxygen partial pressure in kinetic interpretation of his own data but his data were inadequate to confirm this assumption. Szuba, et al.(11) report the oxidation to be second order with respect to oxygen. They investigated the oxidation of Silesian hard coals with air over the range of 340 to 520°F. Gaberman and Khalitov(12) oxidized coal with air in a fluid bed up to 15 atm at about 350°F. They found the reaction to be markedly faster under pressure.

It is noted that when the reaction is diffusion controlled, the rate will be first order with respect to oxygen concentration. Many experimenters varied inlet oxygen partial pressure, but operated with zero exit concentration, and this made evaluation of kinetic effects impossible.

d. Effect of Temperature

Literature values for the activation energy of the coal-oxygen reaction vary from 9 to 35 Kcal/mole. Struck(5) reports the data of Sebastian and Mayers(13) as given below:

Lignites and low-rank bituminous	12,000 cal/mole O_2
High-rank bituminous and anthracite	18,000 cal/mole O_2
High temperature coke	30,000 cal/mole O_2

Struck's own values(6) were about 12 Kcal/mole for Pittsburgh seam coal and 33 Kcal/mole for Disco char. Szuba, et al.(14) report an activation energy of 9.3 Kcal/mole for Silesian hard coals.

A problem associated with evaluation of temperature effects is whether significant diffusion and heat transfer effects are present. Diffusion effects tend to lower the activation energy. Heat transfer limitations raise the internal solid temperature above the nominal system temperature.

e. Effect of Particle Size - Diffusion

Struck(6) found that at 650°F, reactivity tended to increase as particle size was reduced. For Montour No. 10 coal, the 100 to 200 mesh material reacted five times as fast as the 35 to 65 and 65 to 150 mesh sizes. For Arkwright coal, the 65 to 150 mesh material was noticeably more reactive than the 35 to 65 mesh coal. Szuba, et al.(11) attributed reduced activation energy in the range of 340-520°F to diffusion effects. Adrzejak and Ignasiak(15) reported that for 0.06-1.5 mm particles at 270°F, oxygen diffusion is controlling for bituminous and anthracite coals but not for brown coals. Peters and Juentgen(16) claimed that activated diffusion of oxygen into the coal is the rate determining step in the oxidation process. Jenkins(17) found the rate of oxygen consumption at about 700°F to be roughly proportional to the external surface area of the coal particle.

f. Heat of Reaction

Jenkins(17) reported the heat of reaction to be 87,000 Btu/lb mole of oxygen. Struck(6) reported values of 181,000 and 190,000 Btu/lb mole of oxygen evaluated at 650°F. Curran, et al.(18) estimated the heat of reaction to be 205,000 to 256,000 Btu/lb mole of oxygen reacted at 500°F. These latter two values are not necessarily inconsistent since Struck dealt with a bituminous coal and Curran studied a subbituminous coal.

Lee, et al.(19) reported on a calorimetric study of the heat of reaction of a Pittsburgh seam coal at 700°F and 800°F under pressure. Their data are presented as a function of the weight loss of the coal; the amount of oxygen consumed is not given. They noted that at 800°F, the rate of devolatilization may compete with the rate of oxidation.

g. Effect of Coal Source

The reactivity of coal generally varies greatly even among coals of the same general type. Struck(6) found marked differences in reactivity between Montour No. 10 and Arkwright coals even though both are Pittsburgh seam coals. The large variation of activation energy among coals has already been discussed. As a broad generalization, reactivity decreases with increasing rank of the coal.

h. Physical Changes Occurring During Preoxidation

When Pittsburgh seam coal is heated over about 700°F, swelling takes place. Mason(20) and Mason and Schora(21) describe the effect of preoxidation at about 750°F as "inflation of the particles to cenospheres ... development of a high-reflectance skin differing in thickness from particle to particle." Forney, et al.(22) reported a volume increase of 100% for 10 to 14 mesh particles and 40% for 48 x 100 mesh particles, both preoxidized at 750°F. The smaller particles swelled only 10% when the pressure was increased to 300 psig. Struck(6) reports results showing that the bulk density of preoxidized coal decreased with preoxidation temperature, i.e., from about 0.55 gm/cc after 700°F treatment to only about 0.25 gm/cc after 800°F preoxidation.

2. Use of Preoxidation to Inhibit Agglomeration

a. Mechanism of Agglomeration

Hiteshue, et al.(23) felt that high molecular weight oils melted and wet the coal. Subsequent cracking forms a sticky residue which can glue particles together. Mason(20) claimed that when preoxidized coal agglomerated, it was due to discharge of material from the interior of the particle and from some particles that may have escaped pretreatment.

It is sufficient for only a small fraction of a charge to be agglomerating in order to make a reactor inoperable. Struck(6) used as little as 25% pretreated material mixed with coke in order to maintain operability in a carbonization step. Experience gained during development of the CO_2 Acceptor Process(2) shows that 18% bituminous coal mixed with non-caking lignite could cause agglomeration in the gasifier.

Forney, et al.(22) showed that agglomeration would occur if helium, steam, or prepurified nitrogen were used to replace a gas stream containing only 0.3% oxygen which had successfully destroyed the agglomerating tendency of the coal.

The Bureau of Mines(24) reported that oxidation produces a skin on the surface of a coal particle which is not wetted by sticky material in the coal.

b. Effect of Preoxidation on Agglomeration

Struck(6) used operability of the Curran carbonization assay as a criterion of successful preoxidation. The assay procedure is to raise the temperature of a fluidized sample at 3°F/min. up to 932°F. He found that at 785°F, 0.6% oxidation was required for Montour No. 10 and 8.5% for Arkwright coal. The operability criterion is whether dilution of the feed coal with char is required to prevent agglomeration and defluidization.

Gasior, et al.(25) treated Pittsburgh seam coal in a fixed bed from 80 to 200 minutes at 660 to 950°F. The char did not agglomerate at 750°F and 3,000 psig in the presence of hydrogen. Forney, et al.(22) used criteria of a free swelling index (FSI) of less than two and no caking in the presence of hydrogen at 1100°F or 1650°F. His experiments were conducted on Pittsburgh seam coals. Successful preoxidations were obtained using a gas containing 0.2% oxygen at about 775°F with a residence time of 5 minutes in a batch fluidized bed reactor. However, the same coal required 2.3% oxygen, 20 minutes and 820°F in a continuous feed reactor.

Using the same criterion for successful preoxidation as Forney, Kavlick and Lee(26) found they required 725-750°F, an oxidation level of 1.0-1.5 SCF O_2/lb of coal, and a residence time of 1-2 hours. Forney, et al. and Kavlick and Lee remarked that their fluid bed products contained particles which had undergone relatively little treatment.

Another technique for effecting solid-gas contact is via entrained flow or free fall. Steinmueller(27) reports that pulverized coal containing 26% volatile matter can be made non-agglomerating by preoxidation while entrained in oxygen containing gas at 700-750°F for 1-3 seconds.

National Fuels Corporation(28) used an entrained coal reactor to preoxidize caking coals prior to briquetting. Coal of -28 mesh was entrained by 5-15 pounds of air per pound of coal for 2 to 4 seconds at 570 to 885°F. Various pilot plants (more than 5 TPD) were operated from 1938 through 1951.

Gasior, et al.(29) preoxidized Pittsburgh seam coal in a free fall system using 5.5-12.7% oxygen, two seconds contact time, 300 psig pressure, and 1040-1250°F. The coal expanded and became sticky during the treatment. Decaking was considered successful if coking did not occur upon exposure to hydrogen at about 1100°F for five minutes.

Forney, et al.(30) describe a staged system operating at about 600 psig. The coal is first pretreated with 0.3-0.8 SCF O_2/lb coal in a free fall preheat section at 752°F for about one second. It is then treated in a carbonizing section where an additional 4 SCF O_2/lb of coal further oxidizes the coal. The final gasification is carried out at 1742°F.

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